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Control of Hydrocarbon Emissions from Cotton and Synthetic Textile Finishing Plants

Foster-Miller Associates, Inc., Waltham, MA

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CONTROL OF HYDROCARBON EMISSIONS FROM COTTON AND SYNTHETIC TEXTILE FINISHING PLANTS

May 1983

by

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16. ABSTRACT The report gives results of an evaluation of the applicability and economics of emissions control technologies for the abatement of volatile organic compounds emanating from cotton and synthetic textile finishing plants. A survey of the state-of-the-art and control technologies design and costing preceded the evaluation. The economic feasibility was determined in two steps: (1) preliminary design, costing, and relative ranking of all identified applicable technologies; and (2) more detailed design, costing, and evaluation of the most economically feasible technologies. A simple payback period approach was taken in the preliminary economic evaluation of the applicable control technologies. Rates of return on capital investment were determined for the final detailed evaluation. Capital and operating costs are provided, allowing interested parties to conduct inhouse evaluations. Carbon bed adsorption with solvent recovery has been identified as the most viable of all technologies, and fluidized-bed carbon adsorption has the best potential to suit the variable operating conditions encountered in textile manufacturing. Potential cost benefits appear attractive.

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ABSTRACT

Foster-Miller, Inc. was contracted by U.S. Environmental Protection Agency (EPA) to evaluate the applicability and economics of emissions control technologies for the abatement of volatile organic compounds emanating from cotton and synthetic textile finishing plants. This report presents the approach and the conclusions resulting from the study. A survey of the state-of-the-art and control technologies design and costing preceded the evaluation. The economic feasibility itself was conducted in two steps: preliminary design, costing and relative ranking of all identified applicable technologies followed by more detailed design, costing and evaluation of the most economically feasible technologies.

A simple payback period approach was taken in the preliminary economic evaluation of the applicable control technologies. Rates of return on capital investment were determined for the final detailed evaluation. Capital and operating costs have been provided to allow interested parties to conduct in-house evaluations.

Carbon bed adsorption with solvent recovery has been identified as the most viable of all technologies and fluidized bed carbon adsorption has the best potential to suit the variable operating conditions encountered in textile manufacturing.

The potential cost benefits, even under far more stringent control requirements than existing regulations for the industry, appear attractive.

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SECTION 1

INTRODUCTION

The United States Environmental Protection Agency (EPA), in promulgating pollution codes and providing guidelines and technical assistance to state, local and other regulatory agencies, conducts state-of-the-art surveys and demonstration programs. These establish the need for control and identify appropriate control technologies to assist them in determining Best Available Techology (BAT), Best Available Technology Economically Achievable (BATEA) and Reasonably Available Control Technology (RACT).

1.1 BACKGROUND

The EPA surveyed and reported (1-3) the significance and impact of hydrocarbon emissions* from textile operations. The survey showed:

- That the processing of textile materials results in the discharge of significant quantities of hydrocarbons to the atmosphere; for example, the contributions to atmospheric hydrocarbon concentrations from all stationary sources by textile operations are 14, 10 and 7 percent in South Carolina, Rhode Island and North Carolina, respectively;

- That the bulk of the hydrocarbons (84 percent) from textile finishing operations, are generated by the drying and heat curing of printed fabrics;

- That, in the year 1986, a potential for hydrocarbon emissions reductions of 42.6 to 403.4, 78.8 to 258.3 and 42 to 157.7 million lb/year in the printing, dyeing and final finishing operations. respectively, exists (4). These figures are projected on the basis of controlling 90 percent of VOC from new or modified equipment only, through the promulgation of new standards in 1981.

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*In this report the terms hydrocarbon, volatile organic compound (VOC) and solvent have been used interchangeably.

1.2 NEED FOR TEXTILE PLANT HYDROCARBON EMISSIONS CONTROL

Hydrocarbons used in the textile industry are derived from liquid fossil fuels. They are expensive and have a high heating value. Thus if the emitted hydrocarbons can be recovered in a pure form for reuse or if their heating value can be utilized in any way, the United States stands to reduce its use of fossil fuel resources.

Through complex atmospheric reactions hydrocarbons form photochemical oxidants which are considered undesirable (5). The significant quantities of hydrocarbons generated suggest that control methods are needed.

1.3 OBJECTIVES

The ultimate objective of the EPA program is the construction and demonstration of a full-scale process for controlling hydrocarbon emissions in the textile finishing industry. The objective of the current Foster-Miller program is to acquire the technical information necessary to determine the appropriateness of control technologies in the textile industry and to design such systems, as well as to develop the economic information necessary to evaluate the viability of the selected control systems.

The technologies considered include those used for hydrocarbon control in other industries. Their appropriateness for the textile finishing industry is determined by the nature of VOC and the mechanism of emissions control.

Several technologies capable of 95 percent control are to be designed in detail; capital and operating costs for them are also developed.

1.4 APPROACH

To ensure the sucess of the current Foster-Miller study, the following scope of work was developed:

- Survey published literature and industry sources to determine "industry-average" nature and extent of emissions from textile fabric finishing operations
- Survey published literature and industry sources for hydrocarbon control practices
- Select technologies capable of 95 percent control when applied to the textile emissions sources

- Produce preliminary design and costs for the selected technologies at industry average stack conditions and a range of emissions concentrations reflecting the variability of the quantity of solvent used

- Conduct an economic evaluation taking into account the seasonal nature of the products of the textile industry and rank the technologies in order of their economic viability

- Based on the preliminary designs and analyses select, design in more detail, and generate costs for two of these technologies

- Project the economics of the technolgies for a range of stack gas flow rates.

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SECTION 2

CONCLUSIONS

The state-of-the-art survey led to the following conclusions:

- Textile fabric production is geared to the marketplace. Volume of products, types of fabric, types of operation (dyeing, printing or final finishing), volume and types of solvent used and hence, the nature and range of emissions, are highly variable.
- Fabric printing has the most potential for high hydrocarbon emissions, and control technologies should address this operation.
- Current emissions regulations are qualitative and address opacity or odor threshold only. Regulations vary from state to state.
- Opacity control is achieved by limiting aerosol emissions, the cause of opacity. Control devices include coalescing filters, electrostatic precipitators (ESPs), direct contact condensers and demisters.
- Odor control is achieved by dilution or incineration.

The preliminary and more detailed design and cost evaluations assumed that any hydrocarbon recovered could be reused or could be used as a fuel. These evaluations resulted in the following conclusions:

- Activated carbon adsorption and incineration are the only technologies capable of achieving over 95 percent emissions control.
- Direct or indirect condensation using refrigeration can approach 95 percent control.
- Activated carbon adsorption projects better economic viability over condensation/aerosol capture technique and various incineration processes over the range of emissions considered.

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- Fluidized bed carbon adsorption shows better economic feasibility than fixed bed carbon adsorption especially when the recovered solvent is used only as fuel.

- Applicable technologies, for 95 percent hydrocarbon removal, rank in the following general order of economic viability on the basis of return on investment:

  - Fluidized bed activated carbon adsorption with recovered solvent reused
  - Fixed bed activated carbon adsorption with recovered solvent reused
  - Fluidized bed activated carbon adsorption with recovered solvent used as fuel
  - Thermal incineration with heat recovery
  - Refrigeration condensation/aerosol capture with recovered solvent reused
  - Fluidized bed catalytic incineration with heat recovery
  - Fixed bed catalytic incineration with heat recovery
  - Fixed bed activated carbon adsorption with recovered solvent used as fuel
  - Refrigeration condensation/aerosol capture with recovered solvent used as fuel.

- Cyclones, scrubbers, fabric filters, demisters, electrically augmented precipitators and other types of particulate (aerosol) collection devices are not suitable for 95 percent control without refrigeration.

- For the best six control technologies and a stack gas concentration of 8000 ppm the ROI's for 1,000 and 10,000 scfm gas flow rates are, respectively, approximately 0.52 and 1.29 times that for a 5,000 scfm system. The changes in ROI with flow rates vary widely when the stack gas concentration is 3000 ppm. At 200 ppm only the fluidized bed carbon adsorption system projects a return and only for flow rates greater than 5000 scfm.

SECTION 3

RECOMMENDATIONS

Recommendations are as follows:

- Solicit the cooperation of a textile finishing plant which has annual solvent usage in their printing operations representative of the industry average; select the most appropriate stack for control.
- Determine stack gas conditions: species emitted, concentration and its variability, flow rate, temperature, moisture content, etc.
- Provide detailed design and cost information for two control technologies suitable for the selected stack.
- Install and evaluate the performance of the more economical control technology.

SECTION 4

LITERATURE AND INDUSTRY SURVEY

The objectives of this effort were:

- To survey published literature and establish emissions characteristics of textile finishing plants and performance characteristics of volatile organic compound (VOC) control systems
- To survey the textile finishing industry to establish industry practices, their effects on emissions and the performance of control technologies in place
- To survey the pollution control equipment manufacturing industry to prepare a list of control systems used for hydrocarbon pollution abatement.

4.1 LITERATURE SURVEY

Foster-Miller's principal sources of published information were journals, industry magazines, EPA reports and books. All the publications that were consulted are listed in Appendix A.

4.1.1 Published Emissions Data

Most articles on emissions from textile finishing sources contained only qualitative information. They provided flowsheets of the different types of operations, for example, dyeing, printing, fabric preparation, curing and heat setting. Some of the more commonly used solvents were listed by generic or trade names. Neither the quantities used nor definitive predictions of the fraction emitted in the exhaust gases, as against the part leaving with the water used, were made available. A few studies quantified the use of organic solvents in the textile industry nationwide or by individual state. In general, the textile plant emissions information obtained from literature was either inadequate for or not applicable to this study. The following qualitative conclusions could be reached from this part of the survey:

- The average quantity of solvent used per pound of fabric is highest for printing as compared with dyeing or final finishing.

- Mineral spirits is the most widely used solvent, especially in printing

- The major part of the solvents used in printing is picked up by the fabric and is ultimately discharged with the exhaust gases

- Whether the exhaust gases are cooled naturally by the atmosphere or quenched artificially, only a small portion of the VOC's is condensed as aerosols

- Most emissions concerns are centered around a blue haze caused by the aerosols mentioned above. The severity of the haze is assessed using Ringelman opacity charts.

4.1.2 Published Control Information

Four generic technologies emerged as the commonly used methods for control of hydrocarbons. The technologies and typical devices in each category are listed below.

- Condensation and aerosol removal--Condensation and aerosol removal technologies include:

    - Condensation:

        - Atmospheric cooling
        - Direct quenching with water spray or scrubbing
        - Indirect cooling with water using heat exchangers
        - Indirect cooling with refrigeration.

    - Aerosol Removal:

        - Coalescing filters
        - Scrubbers (several types)
        - Cyclone followed by scrubbers
        - Demisters (several types)
        - Wet electrostatic precipitators
        - Mechanically assisted cyclones.

- Incineration--Incineration is commonly accomplished by:

    - Direct flame incineration
    - Fixed bed catalytic incineration
    - Fluidized bed catalytic incineration.

- Adsorption--The two adsorbents commonly used for hydrocarbon control are activated carbon and molecular sieves. These can be used in:
    - Fixed bed adsorption
    - Fluidized bed adsorption.
- Absorption--Absorption technology uses:
    - Absorbents selected to match absorbates. Acetone/Water, Benzene/Kerosene, Methyl-Ethyl Ketone/Water are examples of typical pairs for gas/liquid absorption systems.

4.2 INDUSTRY SURVEY

The data base used in this study to define the characteristics of a typical textile finishing plant in the United States was developed from the following three sources:

- Foster-Miller/American Textile Manufacturers Institute (ATMI) Survey, conducted as a part of this study
- OAQPS Survey (unpublished)
- Monsanto Emission Measurements (unpublished).

Each of these sources is discussed briefly below.

4.2.1 Foster-Miller/American Textile Manufacturers Institute Survey

Foster-Miller sent out a questionnaire (see Appendix B) on behalf of the ATMI to 250 textile companies in the United States from which 21 responses were received. Of these only 15 reported production data. Further clarification of the data and procurement of additional information was carried out with ATMI's assistance.

The total processing capacity of broadwoven and knit fabrics was available for 1977 (4). Based on the past 3 years' data, it was assumed that the total capacity in 1978 was the same as in 1977. As shown in Table 1, the 15 companies responding with capacity information represent 6.0 percent of the total number and possess 11 percent of the total capacity in the United States.

TABLE 1. TOTAL CAPACITY OF TEXTILE FINISHING COMPANIES WHICH RESPONDED TO FOSTER-MILLER QUESTIONNAIRE

Textile Finishing Companies	Number	Percent of Total	Total Capacity in Million Pounds (1978)	Percent of Total Capacity
Responding to questionnaire	15	6	599	11
Total in United States	250	100	5320	100

4.2.2 EPA Data

EPA/OAQPS Data--

Two responses which included capacity and/or type of material processed were received from Mr. Dennis Crumpler of OAQPS. These have been coded as number 22 and 23 as shown in Table 2. Additionally, we received data primarily relating to the emissions from nine printing plants from Mr. Greg Gaspercz of OAQPS. As shown in the table, these have been coded from 24 through 32.

Monsanto Data--

Applicable Monsanto emissions data on four plants have also been included in the data base. These plants have been coded 33 through 36 as shown in Table 2.

4.2.3 Control Devices Used in Industry

In the absence of stringent regulations to control organic emissions, textile plants, in general, have not made a concerted effort to do so. However, the enforcement of other laws curbing opacity, odor emission, and general nuisance, have led to abatement of particulates (condensed aerosols, smoke, etc.) issuing from stacks. Coalescer/demisters, electrostatic precipitators and scrubbers are used to control aerosols and smoke.

A few textile plants use flame incinerators, catalytic incinerators and carbon bed adsorbers to achieve a higher degree of organic emissions control with recovery of heat from incinerators and usable solvents from carbon beds (1, 3, 6). Figure 1 shows the distribution of control equipment in the textile industry by type of device.

TABLE 2. EXPLANATION OF CODE NUMBERS AND DATA SOURCES

Code Number	Data Source	
1 through 21	Foster-Miller Questionnaire	
22 A, B	Plant AA - OAQPS	Unpublished
23	Plant BB - OAQPS	
24	Plant A - OAQPS	
25	Plant B - OAQPS	
26	Plant C - OAQPS	
27	Plant D - OAQPS	
28	Plant E - OAQPS	
29	Plant F - OAQPS	
30	Plant G - OAQPS	
31	Plant H - OAQPS	
32	Plant I - OAQPS	
33	Plant A - Monsanto	Unpublished
34	Plant B - Monsanto	
35	Plant C - Monsanto	
36	Plant D - Monsanto	

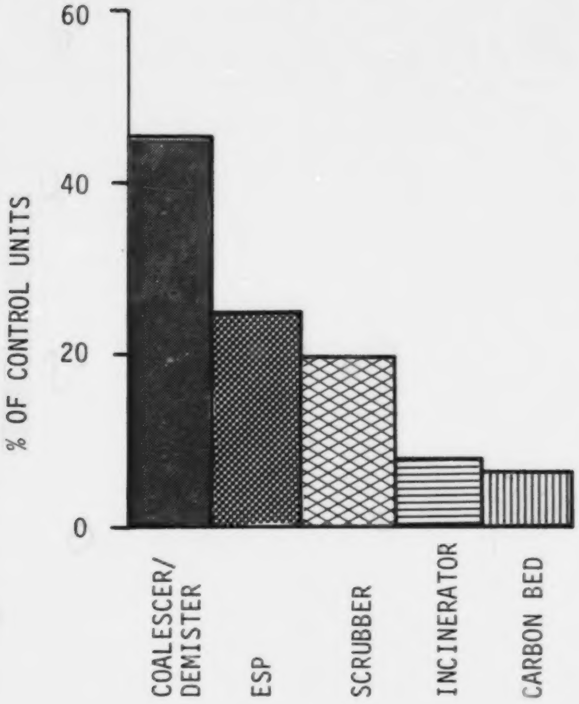

Figure 1. Distribution of control equipment by type.

Published information on effectiveness for any of these devices was not found. Ninety to 98 percent capture by electrostatic precipitators is claimed by manufacturers. These figures must be based on particulate concentration and not total organic emissions. Since most devices were installed to reduce opacity, the control efficiency is judged qualitatively by eye, sometimes using Ringelman charts.

A list of control devices and manufacturers appears in Appendix C.

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SECTION 5

DETERMINATION OF AVERAGE TEXTILE INDUSTRY PRACTICES AND SELECTION OF DESIGN PARAMETERS

5.1 PRODUCTION RATES

5.1.1 Overall Production

The annual fabric production and type of fabric of the textile finishing plants for which this information was available is given in Table 3. The methodology used for the selection of typical capacity for use in this study is explained below.

Plant capacities are plotted on a log-log scale to illustrate the relative density of plants with respect to the order of magnitude of their production capacity. The capacity range chosen as representative of industrial practice is based on the arithmetic average determined from this study for each group of plants processing either cotton, synthetics or blends. The typical range for all plants collectively is based on the arithmetic average for the U.S. (4). The range represents plant capacities of the same order of magnitude as the mean capacity for the group. This is determined by the log value of the plant capacity and is limited to values approximately 25 percent higher and lower than the log of the mean plant capacity. This range includes 30 to 50 percent of the plants for which production data was available.

Figures 2, 3, 4 and 5 show plant capacities for all fabrics, cotton, synthetics, and blends, respectively. The capacity range assumed to be typical of each category is shown by the shaded portion. Respondents 11 and 22 provided information on more than one plant. Therefore, 11A, 11B, 11C, and 11D and also 22A and 22B have been considered separately.

Typical plant capacities for various fabrics are shown below.

Fabric	Capacity (million lb/year)
All	10 to 50
Cotton	5 to 20
Synthetics	5 to 20
Blends	15 to 100

TABLE 3. ANNUAL FABRIC PRODUCTION OF TEXTILE FINISHING PLANTS PROCESSING COTTON, SYNTHETICS OR BLENDS

(Figures in million lb/year)

Code number	Cotton	Synthetics	Blends	Total mix	Remarks
1		X		X	
2		9.1		9.1	
3		2.3		2.3	
4	X			X	Narrow fabrics (labels)
5		X		X	
6		5.7	21.6	27.3	
7	X	X		5.0	
8		X		X	
9	X	X	X	X	
10	24.3		54.0	78.3	
11A	12.6	41.1	36.5	90.2	
C	0.1	4.4	2.3	6.5	
D	18.9	2.4	18.0	39.3	
12	15.7		4.4	20.1	
13		9.0		9.0	
14	X	X		X	
15	X	X		0.5	
16	0.3	1.5	3.1	4.9	
17	12.9	25.8	11.3	50.0	
18		11.5		11.5	
19	X	X		15.3	
20		1.8		1.8	
21	X	X		X	
22A	1.9	1.1	175.3	178.3	
B			50.0	50.0	
23		X		X	
	10.8	9.6	37.6	33.3	Arithmetic averages

X - Capacity not reported

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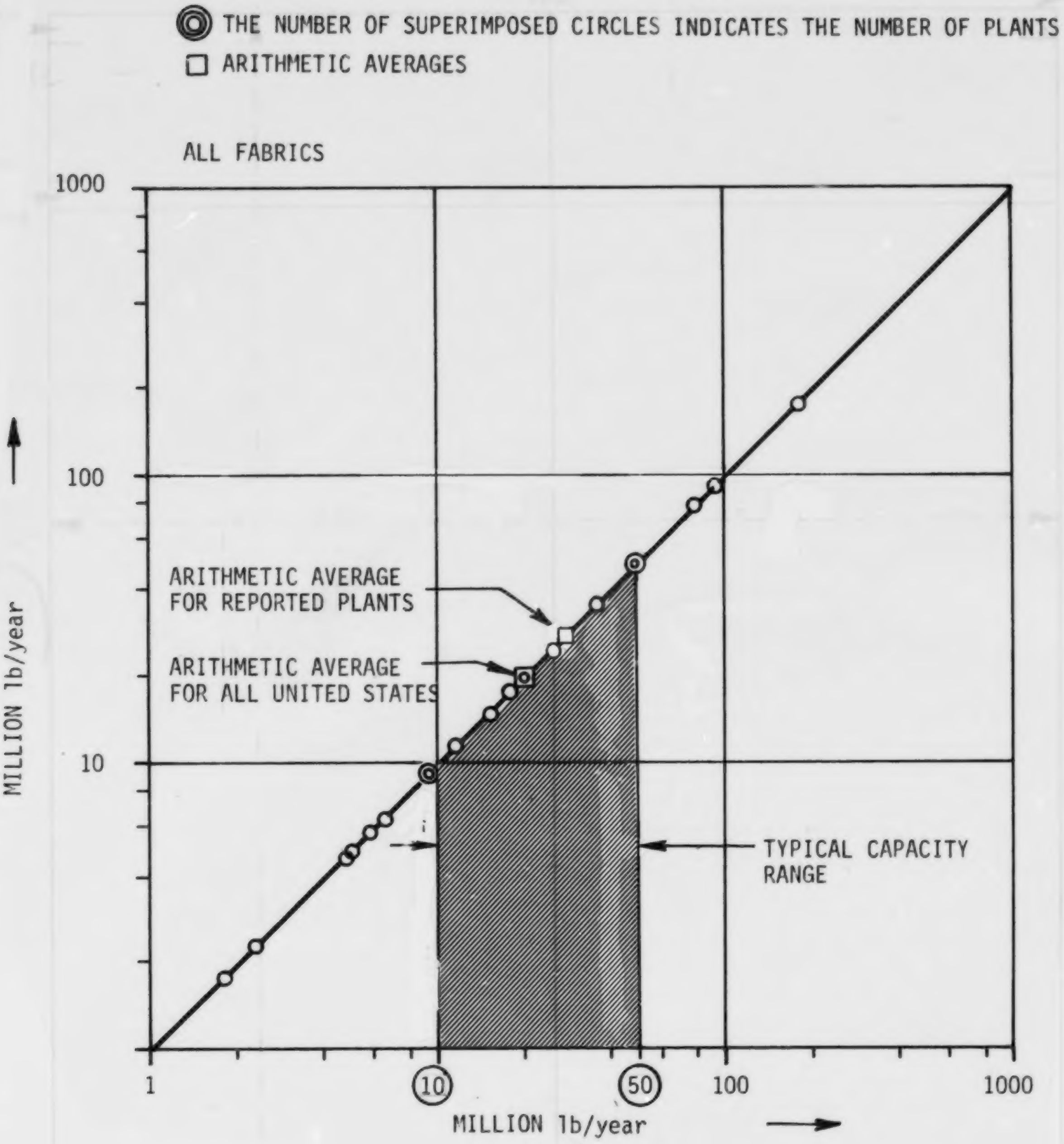

Figure 2. Reported capacities of textile finishing plants processing all types of fabrics.

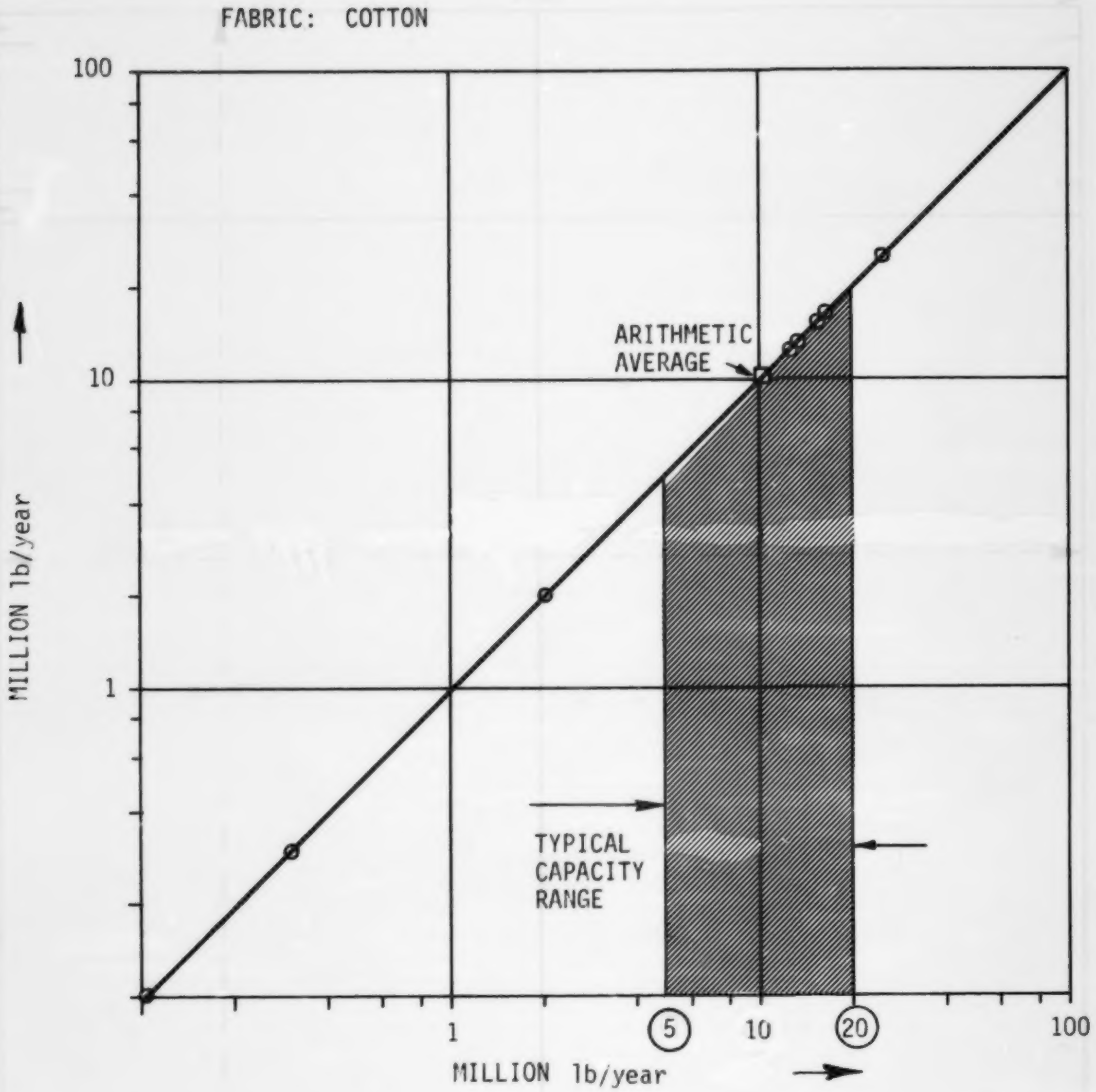

Figure 3. Reported capacities of textile finishing plants processing cotton.

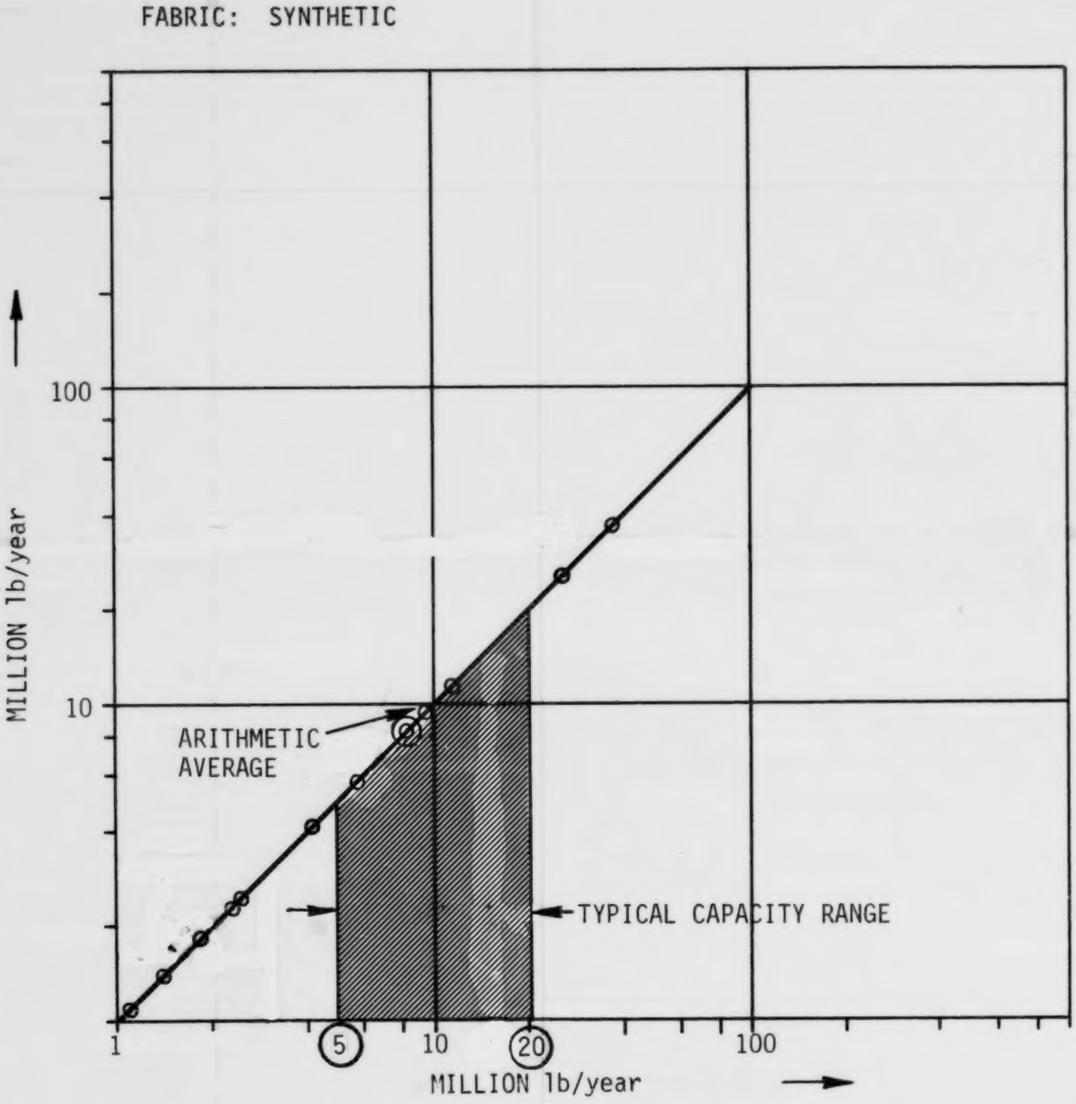

Figure 4. Reported capacities of textile finishing plants processing synthetics.

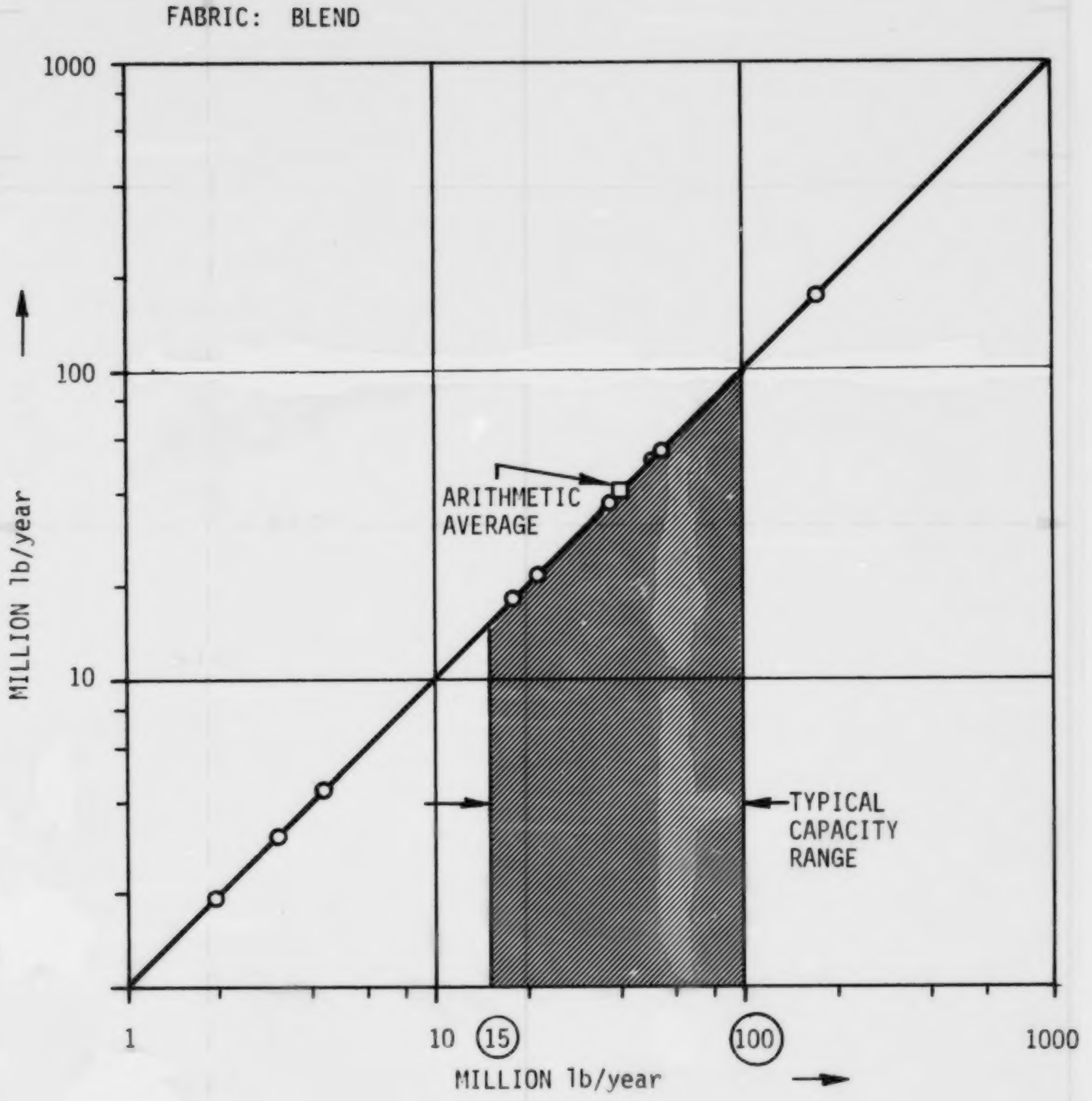

Figure 5. Reported capacities of textile finishing plants processing blends.

5.1.2 Production by Type of Fabric

Among the 26 plants reporting the type of fabric processed, nine processed only synthetic fabrics, one processed only cotton, and one processed only blends, whereas 15 processed some combination of cotton, synthetics and blends. The number of plants processing various combinations of fabrics are given below in decreasing order.

Fabric Processed	Total Plants
Cotton, synthetics and blends	7
Cotton and synthetics	5
Cotton and blends	2
Synthetics and blends	1
Total	15

In the majority of cases, the blends were of the cotton-polyester type.

A typical plant in the United States is, therefore, presumed to process some combination of cotton, synthetics and blends. The typical blends are presumed to be of the polyester-cotton type.

5.1.3 Production by Operations

The total number of finishers in the United States has been estimated at close to 250 (1). The total number of plants having printing operations is estimated to be 115 (4). Thus, a typical plant would be either a dyeing-finishing operation or a printing-finishing operation. The number of plants responding to the Foster-Miller survey which have dyeing and finishing operations is greater than the number of plants reporting printing and finishing operations. Considering the emphasis on printing as a prime source of HC emissions, however, the characteristics of a plant which has each of dyeing, printing and finishing operations would be more representative of the entire finishing industry.

Due to the limited amount of data available, it was not possible to select typical annual process capacity based on a plant which carried out all three fabric operations, i.e. dyeing, printing and finishing. The range of annual production reported in the survey for each operation is given in Table 4 along with operating time and the estimated number of lines of operation.

Although a few plants are known to carry out more final finishing than dyeing or printing, we assume that in a typical dyeing-finishing facility the amount of fabric dyed is the same

TABLE 4. RANGE OF ANNUAL PRODUCTION BY OPERATION

Operation	Annual production (million lb)	Annual operating time (hr)	Number of lines
Heat setting	0.5 - 50	600 - 12,000	1 - 3
Dyeing	2 - 90	5,000 - 30,000	1 - 5
Printing	6 - 40	6,000 - 100,000	1 - 16
Finishing*	2 - 80	4,000 - 30,000	1 - 5

*Functional finish or final finish.

as the amount finished. Also, for a printing-finishing facility, we assume that whatever fabric is printed is subsequently finished. Thus, for a 20 million lb capacity plant, we assume that 10 million is dyed and finished and another 10 million is printed and finished. The typical operating time and number of lines for such a plant are given in Table 5.

TABLE 5. TYPICAL ANNUAL PRODUCTION BY OPERATION

Operation	Typical annual production (million lb)	Typical annual operating time (hr)	Typical number of lines
Heat setting	5	5,000	1
Dyeing	10	10,000	2
Printing	10	20,000	4
Finishing	20	20,000	4

5.2 SOLVENT USE

5.2.1 Fabric Processing Rate

The process characteristics considered in this study include the fabric processing rate, the type of solvent used and solvent usage rate in pounds of solvent per 1000 lb of fabric. The fabric processing rates for heat setting, dyeing, printing and finishing operations as well as the annual operating times are listed in Table 6. The solvent type and usage rate for each operation are presented separately in Tables 7, 8 and 9.

Dyeing Process--

From the emissions characteristics, the plants surveyed could be classified into two groups. The group (Code Nos. 11C, and 18) whose solvent usage was high probably used variations of atmospheric beck processes which need organic carriers. The group with lower usage of organic solvents (Code Nos. 2, 10, 11A, 11D, 13 and 16) must have used either high temperature processes like the Thermosol process or pigment dyeing processes. The trend is towards using high temperature processes.

Typical solvents used seem to be Xylene, Methanol and Perchloroethylene. Typical solvent usage was less than 10 lb/ 1000 lb of fabric processed, as reported in Table 7.

Printing Process--

In the estimation of industry average emissions factor and solvent usage for printing operations an appreciation of the impact of several printing process variables (4) is required. In general terms, these are:

- Dye based print pastes contain more solvent than pigment based pastes.
- Dyes penetrate fabric fibers, whereas pigments are applied to the fabric surface and cured. Thus, when processing different blends of fibers, it is easier to use pigments.
- Screen printing operations tend to use higher solvent pastes than roller printing operations.
- Screen printing produces less defined patterns and is associated with non-apparel fabric production. Apparel fabrics are produced using roller printing. However, special precautions in choosing the method and the paste will allow reversal of practices. For example, a flat screen and high viscosity paste will result in better pattern definition.

TABLE 6. REPORTED AVERAGE FABRIC PROCESSING RATES FOR HEAT SETTING, DYEING, PRINTING, AND FINISHING OPERATIONS

Code Number	Operation	Fabric processing rate per train (yd/min)	Total operating time (hr/yr)	Estimated number of trains	Material	Fabric production (million lb/yr)	Solvent usage
2	Heat setting	40 to 45	-	-	synthetics	-	Yes
3	Heat setting	-	3,872	1	synthetics	-	No
	Dyeing	-	5,808	1		-	No
	Finishing	-	7,600	2		-	No
6	Heat setting	90	735	1	synthetics, blends	1.3	No
	Printing	27	11,322	2		5.9	Yes
	Finishing	53	21,174	3		21.6	No
7	Dyeing	15	4,844	1	cotton, synthetics	5.0	No
	Finishing	17	4,384	1		5.0	Yes
10	Heat setting	97	12,219	3	blends cotton, blends	49.8	No
	Dyeing	93	19,224	3		78.3	Yes
	Finishing	93	19,224	3		78.3	No
11A	Heat setting	78	10,622	2	cotton, blends	36.4	Yes
	Dyeing	68	30,273	5		90.2	Yes
	Finishing	60	28,518	5		74.3	Yes
11D	Heat setting	93	12,241	2	cotton, blends, synthetics	26.1	No
	Dyeing	31	16,772	3		12.1	Yes
	Printing	17	100,031	16		39.3	Yes
	Finishing	50	28,297	5		32.3	Yes

(continued)

TABLE 6. (continued)

Code Number	Operation	Fabric processing rate per train (yd/min)	Total operating time (hr/yr)	Estimated number of trains	Material	Fabric production (million lb/yr)	Solvent usage
12	Heat setting	31	1,468	1	cotton, blends	2.7	Yes
	Dyeing	69	9,795	2		6.6	No
	Finishing	79	15,008	3		11.6	No
13	Heat setting	50	7,344	1	synthetics	7.9	No
14	Heat setting	10	-	-	cotton, synthetics	-	No
	Printing	10	-	-		-	Yes
15	Heat setting	17	634	1	cotton, synthetics	0.5	No
16	Heat setting	20	4,800	1	cotton, synthetics	4.0	No
	Dyeing	18	6,000	1		4.9	Yes
	Finishing	12	4,300	1		2.5	No
18	Finishing	70 to 80	4,200 to 4,800	2	synthetics	7.6	Yes
19	Heat setting	27	954	1	cotton, synthetics	1.1	Yes
	Printing	60	5,760	1		14.3	Yes
	Finishing	60	5,760	1		14.3	No
24	Printing	24	-	-	-	-	Yes
25	Printing	37	-	-	-	-	Yes

(continued)

TABLE 6. (continued)

Code Number	Operation	Fabric processing rate per train (yd/min)	Total operating time (hr/yr)	Estimated number of trains	Material	Fabric production (million lb/yr)	Solvent usage
26	Printing	48	-	-	-	-	Yes
27	Printing	48	-	-	-	-	Yes
29	Printing	54	-	-	-	-	Yes
30	Printing	10	-	-	-	-	Yes
31	Printing	67	-	-	-	-	Yes
32	Printing	19	-	-	-	-	Yes
33	Heat setting	65 to 135	-	-	blends	-	No
	Dyeing	90 to 115	-	-		-	-
	Finishing	45 to 125	-	-		-	-
34	Dyeing	55 to 100	-	-	blends	-	Yes
	Finishing	50 to 60	-	-		-	-
35	Dyeing	60 to 100	-	-	blends	-	-
	Finishing	60 to 80	-	-		-	Yes
36	Finishing	50 to 60	-	-	blends	-	-

TABLE 7. SOLVENT USAGE FOR DYEING OPERATIONS

Code number	Fabric	Solvent type	Annual solvent consumption (lb)	Annual fabric production (million lb)	Pounds solvent/ 1000-lb fabric
1	Synthetics	CT Carrier LS	20,000	-	-
		Palo Carrier 270	20,000	-	-
2	Synthetics	Xylene	42,480	9.1	4.7
8	Synthetics	Varsol	20,000	-	-
10	Cotton, blends	Stoddard solvent	710,000	78.3	9.1
11A	Cotton, synthetics, blends	Formaldehyde-based pigment solvent	38,900	90.2	1.0
		Methanol	21,300		
		Emulsified solvent	21,300		
		Trichlorobenzene/butyl	7,350		
11B	-	Methanol	54,298	-	-
11C	Cotton, synthetics, blends	Methyl Napthalene	192,000	≤6.5	≥29.4
11D	Cotton, synthetics, blends	Isopropanol	1,890	12.1	0.2
13	Synthetics	Perchloroethylene	25,000	≤9.0	≥2.8

(continued)

TABLE 7. (continued)

Code number	Fabric	Solvent type	Annual solvent consumption (lb)	Annual fabric production (million lb)	Pounds solvent/ 1000-lb fabric
16	Cotton, syntnetics, blends	Xylene	40,000	4.86	8.2
18	Synthetics,	Mineral Spirits	500	≤11.5	≥46.6
		Methanol	135,504		
		Xylol	397,943		
		Perchloroethylene	2,200		

TABLE 8. SOLVENT USAGE FOR PRINTING OPERATIONS

Code number	Fabric	Solvent type	Annual solvent consumption (lb)	Annual fabric production (million lb)	Pounds solvent/ 1000-lb fabric
6	Synthetics, blends	Mineral Spirits	18,800	5.9	3.7
		2 Ethyl Hexanol	3,055		
11D	Cotton, synthetics, blends	Varsol	3,934,000	39.3	100.1
14	Cotton, synthetics	Varsol	18,801	-	-
19	Cotton, synthetics	Varsol 1	670,000	14.3	49.1
		Dybln	31,800		
		Xylol	1,000		
22B	Blends	Varsol	369,000	50.0	7.4
24	-	Varsol	-	-	3.0
25	-	Varsol	-	-	8.2
26	-	Varsol	-	-	77.5
27	-	Varsol	-	-	3.0
28	-	Varsol	-	-	80.5
29	-	Varsol	-	-	17.5
30	-	Varsol	-	-	59.0
31	-	Varsol	-	-	0.6
32	-	Varsol	-	-	153.0

TABLE 9. ADDITIVE USAGE FOR FINISHING OPERATIONS

Code number	Fabric	Additive type	Annual solvent consumption (lb)	Annual fabric production (million lb)	Pounds solvent/ 1000-lb fabric
7	Cotton, synthetics	Varsol	121,216	5.0	28.0
		Propanol	18,500		
11A	Cotton, synthetics, blends	Glyoxal-based resin	2,133,380	74.3	45.9
		Melamine-Formaldehyde resin	12,039		
		Acrylic-Self Crosslinking (contains formaldehyde)	1,263,397		
		Pigment binder (contains formaldehyde)	2,550		
11D	Cotton, synthetics, blends	Isopropanol	15,000	32.3	0.5
18	Synthetics	Methanol	167,455	≤11.5	≥14.6

- Printed fabrics are first dried to render the pattern smudge proof and often wet final finished before being cured. Hence, emission factors depend on print paste solvent content, wet pick-up, fabric processing rate, and degree of drying before being cured.

Reported solvent usage rates for printing range from 0.6 to 153 lb/1000 lb of fabric. This wide range of values is attributed to the wide variety of printing techniques available as described above. In half of the plants less than 20 lb of solvent per 1000 lb of fabric was used. The preferred solvent for printing appeared to be Varsol.

Finishing Process Characteristics--

The plants surveyed did not, in the majority of cases, use organic solvents is their finishing operations. The typical finishing formulation when applied consisted of formaldehyde or glyoxal-based resin. The typical resin usage was on the order of 50 lb/1000 lb of fabric processed.

5.2.2 Average Stack Gas Characteristics

The reported stack gas characteristics are listed in Tables 10 through 13 for heat setting, dyeing, finishing and printing operations. The stack gas temperatures depend upon the specific process being vented. The typical stack gas flow rate seems to fall in the range of 4000 to 8000 scfm each for heat setting, dyeing, printing and finishing operations. The number of stacks varies with the capacity of the plant.

5.3 AVERAGE FABRIC PROCESSING PLANT PARAMETERS

The following conclusions were reached as a result of the textile industry survey data base (this study and reference 4):

- Average fabric processing rate: 60 yd/min
- Average fabric weight: 0.5 lb/yd
- Average wet pick-up: 70 percent
- Solvent content of print pastes can be as low as 0.5 percent, or as high as 50 percent, and in some instances can reach 70 percent
- Mineral spirits is the single most commonly-used solvent
- Average exhausts from dryers or tenter frames amount to 5000 scfm
- Nominal stack gas temperature is 300°F.

TABLE 10. REPORTED STACK GAS CONDITIONS FOR A TYPICAL LINE OF TEXTILE HEAT SETTING OPERATIONS

Code number	Number and source of stacks	Average gas flow rate (scfm/stack)	Temperature (^{o}F)	Contaminant
2	1-T*	1,700	350	Mineral oil
	1-T	7,500	265	
	1-T	4,800	310	
6	1	10,000 rated	350	None
10	1-P*	13,700	250	None
	1-T	8,400	375	
11A	3-T	4,800	300	Xylene-based solvent
	1-C*	5,300	320	
11D	5-T	3,300	320	None
15	3-T	4,300	340	None
16	3	7,575	185	None
19	1	3,600	145	Varsol
	1	3,800	140	
	1	4,300	525	
20	3	7,600	225	None

*T - Tenter
P - Preheater
C - Contact

TABLE 11. REPORTED STACK GAS CONDITIONS FOR A TYPICAL LINE OF TEXTILE DYEING OPERATIONS

Code number	Number and source of stacks	Average gas flow rate (scfm/stack)	Temperature (^{O}F)	Contaminant
1	2	2,900	350	CT Carrier LS Palo Carrier 270
5	1-P*	2,400	180	None
10	1-P	26,900	150	Stoddard solvent
	1-Th*	27,700	95	
11A	1-St*	1,600	180	Formaldehyde Methanol Emulsified solvent Trichlorobenzene + Butyl
	1-P	3,900	350	
	1-Th	5,300	325	
	2-W*	6,900	140	
	2-D*	6,500	180	
11D	7	3,700	175	Isopropanol

*P - Predryer
Th - Thermosol
St - Steamer
W - Wash box hood
D - Dry can hood

TABLE 12. REPORTED STACK GAS CONDITIONS FOR A TYPICAL LINE OF TEXTILE FINISHING OPERATIONS

Code number	Number and source of stacks	Gas flow rate (scfm/stack)	Temperature (°F)	Contaminant
1	4	1,900	150	None
5	4-O*	2,400	225	None
6	1	10,000, rated	350	None
9	4	11,000	300	None
10	1-S	5,900	275	None
	1-P	23,600	95	
	2-T	7,000	290	
	1-O	23,600	95	
11A	1-P	3,900	350	Glyoxal and Formaldehyde based resins
	2-T	5,500	290	
	1-O	5,300	325	
11D	3-T	4,500	240	Isopropanol
15	6	6,700	320	-
19	1	3,800	140	None

*P - Predryer; T - Tenter frames; O - Curing oven; S - singer

TABLE 13. REPORTED STACK GAS CONDITIONS FOR A TYPICAL LINE OF TEXTILE PRINTING OPERATIONS

Code number	Number of stacks	Gas flow rate (scfm/stack)	Temperature (°F)	Contaminant
6	2	7,500, rated	275	Mineral Spirits 2 Ethyl Hexanol
11D	2	2,700	300	Varsol
19	1	4,500	125	Varsol 1 Dybln Xylol

5.3.1 Prediction of Emission Concentrations

Due to the high frequency of solvent usage and the solvent usage rate for printing operations, it can be concluded that textile printing operations have the most potential to contribute significantly to plant VOC emissions. Therefore, estimated concentrations of VOC in plant exhaust will be based on typical print plant parameters using mineral spirits as the most widely-used solvent.

Properties of Mineral Spirits--

Mineral spirits contain approximately 40 percent by weight of aliphatic and 40 percent cyclic hydrocarbons. The rest is made up of alkyl benzene. Weight distribution by carbon chain length is given in Figure 6. It can be seen that C_{10} aliphatic and aromatic compounds predominate. The average molecular weight is 140, approximating a $C_{10}H_{20}$ hydrocarbon.

Table 14 lists properties of Varsol, a representative mineral spirit, supplied by manufacturers. Detailed thermodynamic properites of mineral spirits are not available. The properties of N-Decane (molecular weight 142; formula $C_{10}H_{22}$) are therefore used when necessary.

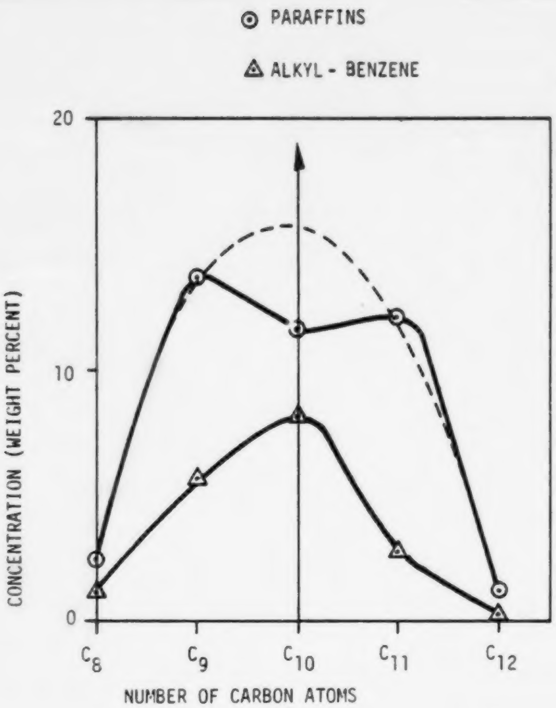

Figure 6. Composition of mineral spirits.

TABLE 14. MANUFACTURER'S DATA FOR VARSOL

Properties important to this study*	Value
Average molecular weight	140
Boiling point range	313 to 403°F
Specific gravity	0.795
Vapor pressure at 100°F	0.7 (psia)
Lower explosive limit	1 (% by volume)
Upper explosive limit	6 (% by volume)
Calorific value	19,930 Btu/lb

*A more comprehensive list of Varsol properties is presented in Appendix D.

Range of Emissions Concentrations--

Emissions concentration is determined from the following:

$$C = 100 \frac{LwWSv}{MQ}$$

where C = Concentration, ppm (vol)

L = Fabric processing rate, yd/min

w = Specific weight of fabric, lb/yd

W = Wet pick-up, weight of print paste absorbed as a percent of fabric weight

S = Solvent content of print paste, percent by weight

M = Average molecular weight of solvent, lb/mole

v = Molal specific volume of solvent, standard cubic feet/mole

Q = Stack gas flow rate, standard cubic feet/minute (scfm).

Given that the average molecular weight of mineral spirits is 140 lb/mole and the molal specific volume is 380 standard cubic feet per mole:

$$C = 270.7 \frac{LwWS}{Q}$$

Hence, for average print line operating conditions, $Q = 5000$, $L = 60$, $W = 0.5$, $W = 70$, concentrations at various solvent contents are as shown in Table 15.

TABLE 15. EXHAUST CONCENTRATIONS AT DIFFERENT PRINT PASTE SOLVENT CONTENTS

S (percent)	C (ppm, volume)
3	340
30	3400
70	8000

5.3.2 Design Emissions Concentrations

Based on the above figures, three emissions concentrations, 200 ppm, 3000 ppm and 8000 ppm, were selected by EPA to cover a range of fabrics, production rates, print pastes and wet pick-up.

5.3.3 Range of Operating Parameters Covered by this Study

The choice of the 200 to 8000 ppm VOC concentration range in the exhaust gases will cover the range of fabric specific weights given in Table 16, for the selected ranges of print paste solvent contents, wet pick-up, and fabric processing rates also shown in Table 16.

TABLE 16. RANGE OF PARAMETERS COVERED BY 200 TO 8000 ppm EMISSIONS CONCENTRATION RANGE

Concentration, C (ppm, vol)	Fabric rate, L (yd/min)	Solvent content, S (%)	Wet pick-up, W (%)	Fabric specific weight, w (lb/yd)
200	20	3	20	3.0
			70	0.9
			100	0.6
		70	20	0.1
			70	-
			100	-
	100	3	20	0.6
			70	0.2
			100	0.1
		70	20	-
			70	-
			100	-
8000	20	3	20	+
			70	+
			100	+
		70	20	5.3
			70	1.5
			100	1.1
	100	3	20	+
			70	7.0
			100	4.9
		70	20	1.1
			70	0.3
			100	0.2

- less than 0.1 lb/yd

+ greater than 10.0 lb/yd

SECTION 6

APPLICABLE CONTROL TECHNOLOGIES

This study calls for the evaluation of control technologies providing 95 percent removal of VOC from the stack gases. The permissible concentrations of VOC in the gases exiting the control device corresponding to the three uncontrolled stack gas concentrations selected for evaluation are given in Table 17.

Generic control technologies for hydrocarbon emissions are:

- Aerosol formation and particulate capture
- Incineration
- Adsorption
- Absorption.

6.1 AEROSOL FORMATION AND PARTICULATE CAPTURE

Gases are capable of carrying varying amounts of hydrocarbons depending on the temperature. The limiting value is the saturation concentration, usually expressed as vapor pressure. Variation of vapor pressure with temperature for ideal gases is given by:

$$\log P = a - \frac{b}{T+c} \qquad (1)$$

TABLE 17. PERMISSIBLE CONTROL DEVICE EXHAUST CONCENTRATIONS FOR 95 PERCENT CONTROL

C Inlet (ppm, volume)	C Exhaust (ppm, volume)
200	10
3000	150
8000	400

where P = vapor pressure

T = temperature

a,b,c = constants for the compound.

Figure 7 shows the saturation concentration of N-Decane at various temperatures at 1 atmosphere total pressure. At any given temperature, excess hydrocarbon (beyond the saturation value) will condense and form aerosol particulates. Aerosol formation is a highly complex mechanism. The size distribution of condensed vapors is not easily predicted, because of its dependence on the nature of the organic compound, presence of nucleating agents such as other aerosols, solid particulates or surfaces, as well as the history of vapor saturation (Example - rate of cooling).

In aerosol formation/particulate capture, the gases are cooled until the saturation vapor concentration is less than the permissible exhaust concentration. The condensed excess vapors (now liquid particulates) are then removed using conventional means.

Conventional particulate control devices are:

- Demisters/cyclones
- Scrubbers
- Filters
- Electrostatic precipitators (ESP).

The first two devices have efficiencies much below 75 percent for submicron particulates with collection efficiency not exceeding 90 percent for the entire particle size range. An overall control of 95 percent is not possible.

Capture efficiencies of filters and ESP's are over 99 and 98 percent, respectively, over submicron and micron particle size range. Severe condensation duty will have to be imposed to achieve overall 95 percent control, as the following calculation illustrates.

Define:

C_i = Inlet VOC concentration, ppm

C_o = Permissible VOC concentration for 95 percent control, ppm

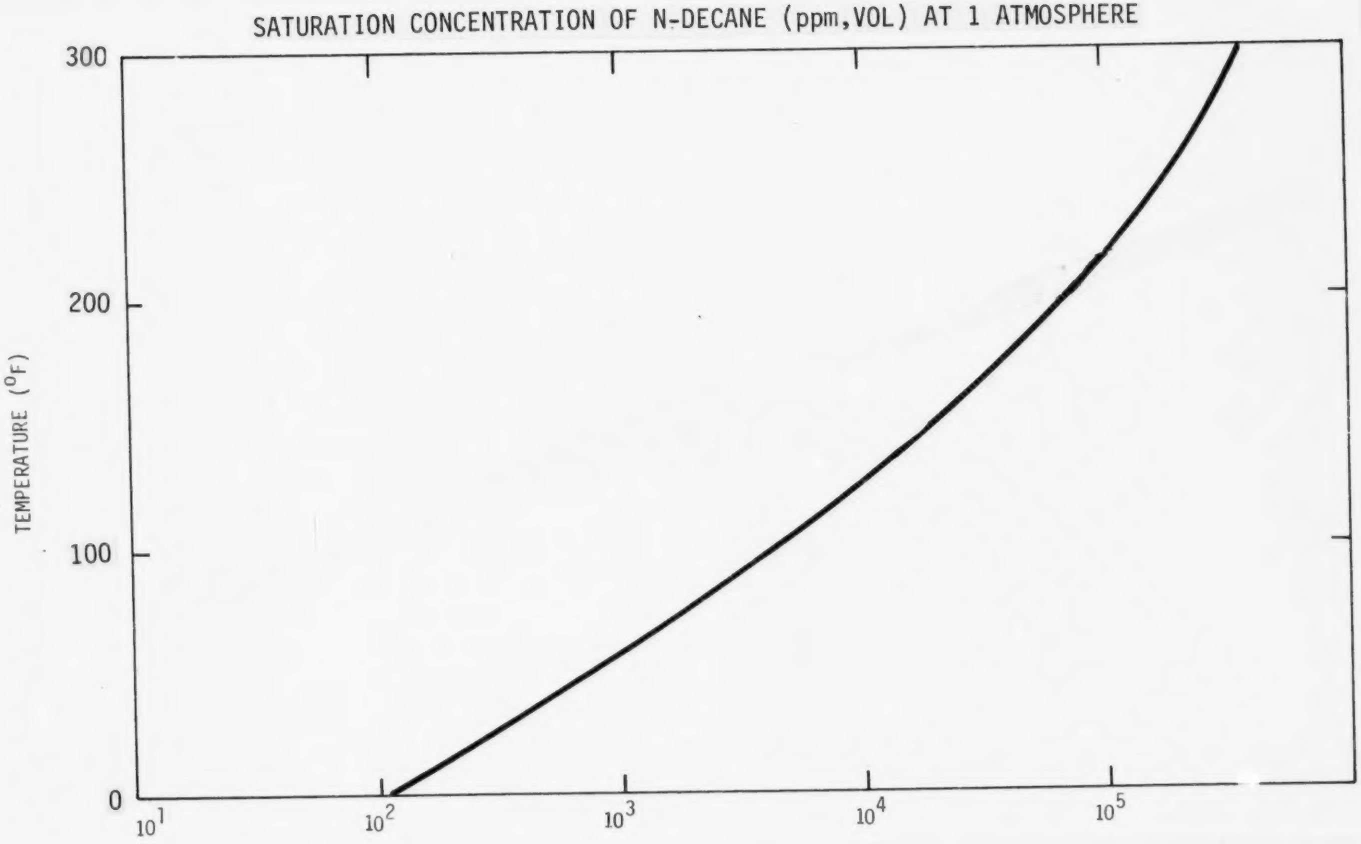

Figure 7. Variation of saturation concentration of N-Decane with temperature.

C_c = Concentration of VOC's which must be condensed, ppm

η_c = Control device efficiency, percent

C_d = Control device exhaust VOC vapor phase concentration, ppm

The concentration remaining in the exhaust after exhaust cooling and particulate removal by ESP will be the sum of the uncondensed VOC's and the condensed VOC's which are not removed by the control device:

$$C_o = (C_i - C_c) + \left(1 - \frac{\eta_c}{100}\right) C_c \tag{2}$$

Thus

$$C_c = 100 \, \frac{(C_i - C_o)}{\eta_c} \tag{3}$$

For

C_i = 3000 ppm

C_o = 150 ppm

η_c = 98 percent

The amount of VOC's which must be condensed is

$$C_c = 100 \, \frac{(3000 - 150)}{98}$$

$$= 2908 \text{ ppm}$$

The maximum gas phase or saturation vapor concentration of VOC in the control device exhaust is then:

$$C_d = C_i - C_c \tag{4}$$

$$= 3000 - 2908$$

$$= 92 \text{ ppm}$$

Thus the vapors will have to be cooled to 0.0°F, as is illustrated in Figure 7. This low temperature can be achieved only with refrigeration.

Hence, refrigeration condensation with filters or ESP is an applicable technology. However, if moisture is present in the gases, which is true for textile dryers, the moisture will condense and freeze in the control device and require defrosting.

6.2 INCINERATION

Incineration involves the burning of all combustible components in the stack gases. The stack gases are generally preheated to the temperature required to support combustion. Modulating controls on the preheater burner regulated by the incinerator discharge temperature are usually used. This allows the fuel gas input to the burner to be reduced as the rate of heat release increases with increasing concentration of combustible vapors.

Incineration can be:

- Thermal
- Catalytic.

In thermal incineration, the auto-ignition temperature is above 1200°F, depending on the nature of the pollutants. For hydrocarbons, 1200°F at a residence of 0.6 sec is successfully practiced in industry for 95 percent control. Higher temperatures can employ shorter residence times (8). The upper temperature limit is set by materials of construction. Thus, when no heat is recovered, a refractory-limited temperature of 2000°F can be reached. This limit is lowered to 1500°F if standard heat exchangers are used for heat recovery. Gases leaving the heat recovery unit are usually not allowed to cool below 300°F to avoid condensation and the resulting corrosion of equipment. Fire codes restrict vapor concentrations to 25 percent of the lower explosive limit. Thirty-five percent of lower explosive concentrations have been used in industry (8). Accordingly, dilution prior to preheating is necessary for concentrations greater than 3500 ppm.

Oxides of aluminum, rare earths or heavy metals such as platinum impregnated in clay, are the more common catalysts for incineration. These minerals or metals are held in a fixed matrix and the gases to be treated are passed through it or, alternatively, the carrier is made in the shape of beads or pellets and the gases to be treated are used to fluidize them. Catalytic incineration takes place at much lower temperatures (500 to 850°F) than thermal incineration. The upper limit is set by the catalyst material and rarely exceeds 1000°F. Hence, when the heating value of the hydrocarbons is low (as in low concentration stack gases), less additional fuel is required to incinerate the VOC's. However, when concentrations are high, more dilution is required to keep the incineration temperature below that which the catalyst will withstand, rendering the equipment larger and more expensive.

A fluidized bed catalytic incinerator permits the introduction of heat exchanger tubes in the bed. Water or air can be passed through the tubes to draw heat from the bed. This flexibility allows incineration of high concentration stack gases without dilution while keeping the catalyst temperature below the upper limit. The heat transfer rate in a fluidized bed is very high and, hence, the required heat exchanger area is relatively small. One drawback is that the equipment size is dictated by the fluidizing velocity which is lower than the velocity in either thermal or fixed matrix incineration. However, the catalysts can be withdrawn for reactivation and charged into the bed without interruption. Thermal, fluidized bed and fixed bed catalytic incineration are applicable control technologies.

6.3 ADSORPTION

Hydrocarbons and other compounds are selectively adsorbed on the surface of such materials as silica gel and activated carbon, as well as natural and synthetic zeolites. Carbon has the most affinity for organic compounds and is widely used for pollution control. The affinity of carbon for VOC's increases with increasing VOC molecular weight, with increasing VOC boiling point, with increasing VOC concentration in the gases, and with decreasing temperature. When VOC (the adsorbate) is adsorbed, it releases heat, thus raising the operating temperature, requiring cooling of the carbon, the adsorbent.

Adsorption is a surface phenomenon and hence, the more surface area occupied by the adsorbate, the more saturated the adsorbent gets, until an equilibrium is reached. This equilibrium is temperature and concentration dependent. Plots of the amount adsorbed versus adsorbate concentration at different temperatures are called adsorption isotherms. Figure 8 shows adsorption isotherms for a carbon/toluene system.

Adsorption is also reversible. By supplying the heat of adsorption, the organic compounds can be desorbed, condensed and collected. Thus, the adsorbate is almost completely recoverable in its original form. This feature makes adsorption/desorption a very attractive control method when the recovered material is valuable. The desorbed carbon can be reused. The process of desorption is called regeneration.

Adsorption can be carried out in batches in fixed beds or continuously in fluidized beds.

In fixed bed adsorption, a vessel is filled with regenerated carbon and the stack gases passed through it. The part of the carbon that first comes in contact with the gases sees the highest concentration of adsorbates and the highest adsorption rate;

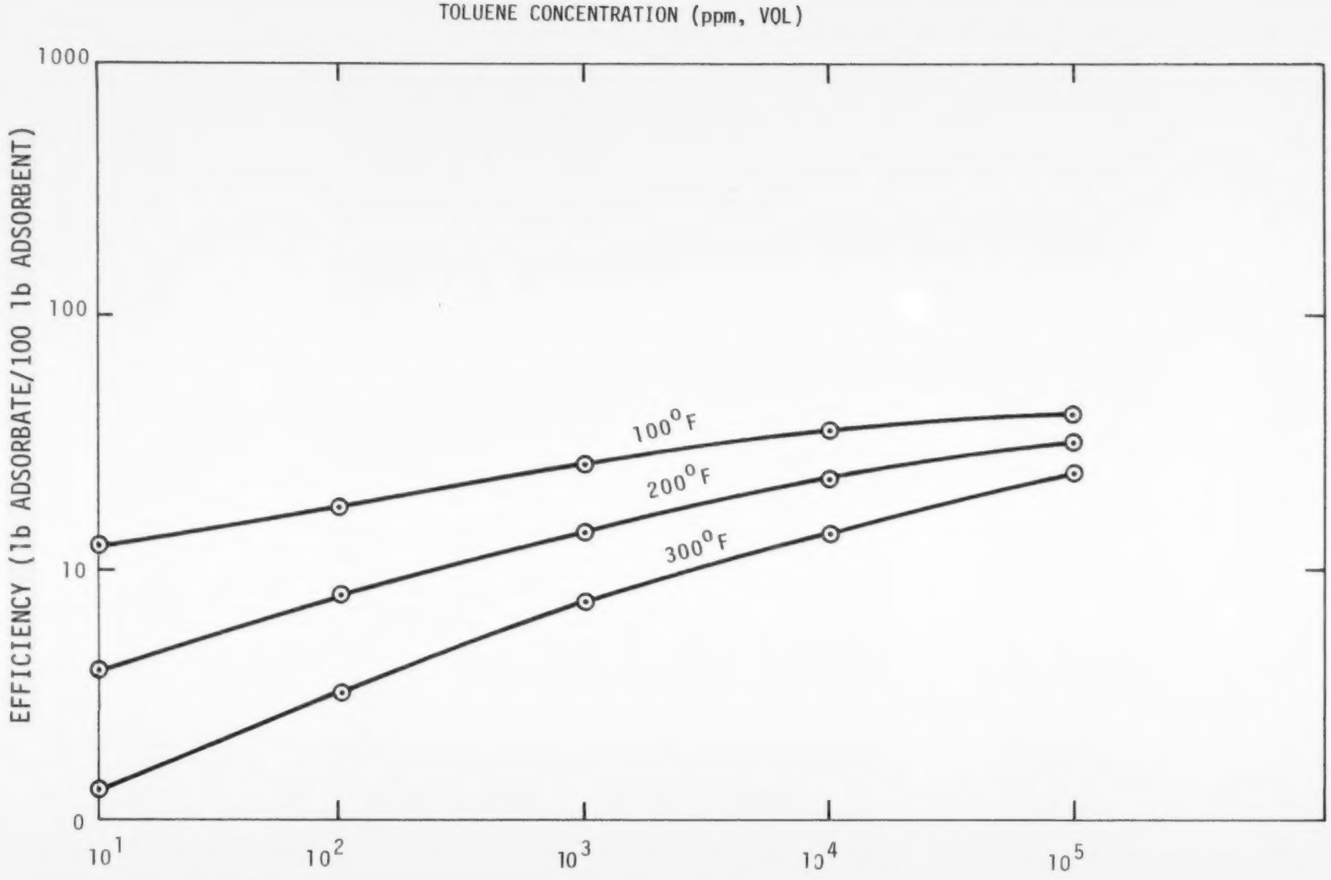

Figure 8. Adsorption isotherms for toluene/activated carbon system.

the "leaner" gases pass through the rest of the bed with decreasing adsorption rate, until the concentrations are so low that no more adsorption can take place. As the carbon at the upstream end gets saturated, the less saturated carbon downstream now sees gases at the highest concentration. In time, the thickness of the saturated layer increases and the remaining bed is unable to adsorb enough VOC and the exhaust concentration exceeds the preset limit. This is called the breakthrough point. At this stage the stack gases are diverted to another vessel with regenerated carbon while the first bed is regenerated. The vessels (beds) are sized so that breakthrough does not occur until the regeneration of the second bed is completed and it is put back in service.

Regeneration is carried out with steam or with heated air or nitrogen. In steam regeneration, live steam is injected into the carbon bed. The heat supplied by the condensing steam desorbs the adsorbate which flows out with the condensate. The two liquids are allowed to separate in a third vessel, from which the solvent is decanted. This method is applicable in cases where the two phases are virtually immiscible. If the solvent tends to mix or react with steam, then the bed is indirectly heated to accomplish the desorption. A recirculating stream of air or nitrogen moves the solvent from the bed to a chiller where the solvent is condensed out and the air or nitrogen is returned to the bed.

In fluidized bed adsorption, the fresh carbon flows down a series of fluidized trays. The exhaust gases enter at the bottom of the absorption section, flow up the trays, and are exhausted from the top tray. Since fresh carbon is introduced at the top and the saturated carbon removed from the bottom continuously, no breakthrough occurs and all the carbon exiting the adsorption section is in dynamic equilibrium with the inlet gases. Maximum carbon utilization is thus achieved.

The carbon leaving the adsorption section falls into the desorption section. This section is a moving packed bed where the carbon is not fluidized. As the carbon passes down, the solvent is desorbed. Regenerated carbon is drawn off at the bottom and conveyed to the top of the adsorption section.

Fluidized bed adsorber size is dictated by the fluidizing velocity, which is low, calling for a large vessel. However, carbon utilization is high and, hence, the operating costs are lower. Both fixed bed and fluidized bed activated carbon adsorption control systems are applicable control technologies for textile plant exhaust emissions control.

6.4 ABSORPTION

Absorption involves the contact of a liquid phase absorbent with the gaseous phase absorbate; either an association or a chemical reaction occurs between the vapors and the liquid. Unlike a scrubbing condenser, where the vapors are first condensed and the resulting aerosol occluded in the liquid phase, the absorbing liquid is always different from the absorbed species. Absorbents for different absorbates have to be selected and optimized. In all cases, to recover the absorbate a more complicated means such as fractionation, distillation or extraction is required. Mineral spirits is a multi-component system and the selection of an optimum absorbent will require in-depth research. Absorption for control of multi-component organic solvents is not currently practiced in industry. There are no known systems offered by equipment manufacturers. Absorption is not considered an applicable technology for textile plant emissions control.

6.5 SUMMARY OF APPLICABLE CONTROL TECHNOLOGIES

The following control systems are capable of removing 95 percent or more hydrocarbons in the concentration range of 200 to 8000 ppm (volume), from textile plant stack gases.

- Refrigeration/condensation + aerosol removal
- Thermal incineration
- Fixed bed catalytic incineration
- Fluidized bed catalytic incineration
- Fixed bed adsorption
- Fluidized bed adsorption.

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- Fixed bed adsorption
- Fluidized bed adsorption.

SECTION 7

PRELIMINARY DESIGN AND EVALUATION OF APPLICABLE CONTROL TECHNOLOGIES

Having identified the control technologies particularly suited for textile printing plant stack gases, a preliminary design and cost assessment of each technology was made using design rules given in engineering manuals. Most of these are based on equipment industry practices and field experience in operating these control devices for VOC pollution abatement both in and outside the textile industry.

7.1 ASSUMPTIONS AND NOMENCLATURE

The following assumptions have been made in the preliminary design and costing of the applicable control technologies. Further assumptions appear under individual system discussions.

Design assumptions are that:

- The design parameters from (8) and (9) are applicable
- Thirty-five percent of lower explosive limit is acceptable (10)
- Gas-to-gas heat exchange temperature approach shall be 50°F or higher (11)
- Gas-to-liquid heat exchange temperature approach shall be approximately 10°F (11)
- Liquid-to-liquid heat exchange temperature approach shall be 5°F or higher (11)
- Gas-to-gas heat transfer coefficient is 3.5 Btu/(hr)(ft^2)(°F)(11)
- Gas-to-liquid heat transfer coefficient is 5.5 Btu/(hr)(ft^2)(°F)(11)
- Liquid-to-liquid heat transfer coefficient is 12 Btu/(hr)(ft^2)(°F)(11)

- Fluidized bed heat transfer coefficient is 35 Btu/(hr)(ft^2)(°F)
- Heat losses and gains are not significant
- Changes in the stack gas physical properties due to control are not significant
- Stack gas flow rate is 5000 scfm
- Stack gas temperature is 300°F.

For 5000 scfm of 300°F stack gas assuming 6 percent relative humidity (RH) air, the perfect gas law was used to determine the mass flow rates of the stack gas components, listed in Table 18.

TABLE 18. MASS FLOW RATES OF STACK GAS COMPONENTS

VOC concentration (ppm, volume)	Mass flow rates (lb/min)			
	Air	VOC	Water vapor	Total
200	278.6	0.4	64.1	343
3000	277.8	5.5	63.9	347
8000	276.4	14.8	63.6	355

Costing assumptions are that:

- Updated equipment costs from (8) and (9) are applicable
- Installation cost varies with the system
- Maintenance cost varies with the system
- Capital cost adjustment factor = (A × B × C) - 1 = 0.2947 where A = 1.07 for tax and freight, B = 1.1 for engineering, and C = 1.1 for contingency
- Heat cost: \$2.9/10^6 Btu (13)
- Electricity cost: \$0.055/kWhr (14)
- Mineral spirits cost: \$0.18/lb (15)
- Labor cost: \$9/hr
- Catalyst regeneration cost: \$5/lb (16)

- Activated carbon cost: \$12/lb (17)
- Equipment operates 6000 hr/year.

The following symbols will be used without subscript to denote stack gas properties:

Q - volume flow rate, scfm

$\dot{m}$ - mass flow rate, lb/min

c_p - specific heat, Btu/(lb)(°F)

h_{fg} - heat of condensation, Btu/lb

H - thermal load, Btu/min

T - temperature, °F

ρ - density, lb/ft^3

and with subscripts:

a - dry air

f - fuel

s - solvent

w - water

to specify individual component rather than mixture properties. Other specific subscripts are used as required.

7.2 REFRIGERATION/CONDENSATION AEROSOL REMOVAL SYSTEM

7.2.1 System Configuration

Figure 9 on the following page shows the system configuration used in the preliminary design.

7.2.2 Preliminary System Design (3000 ppm case)

System temperatures illustrated in Figure 9 are determined according to the assumptions listed in subsection 7.1 and derived in the following calculations.

Condenser Outlet Temperature--
From Equations 3 and 4 a condenser outlet concentration of 92 ppm will be required for a stack gas concentration of 3000 ppm

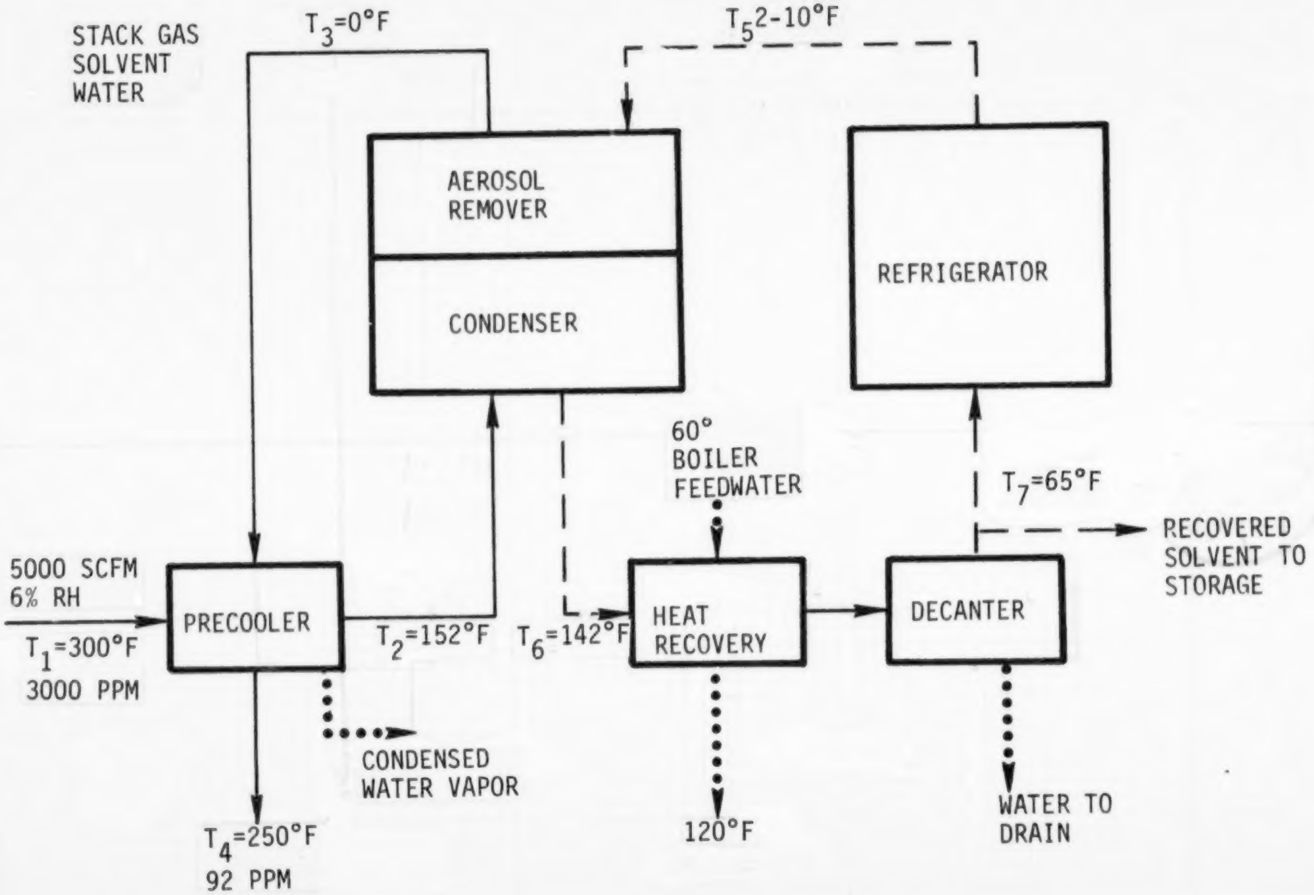
STACK GAS
SOLVENT
WATER
T_3=0°F
$T_5$2-10°F
AEROSOL REMOVER
CONDENSER
REFRIGERATOR
5000 SCFM
6% RH
T_1=300°F
3000 PPM
PRECOOLER
T_2=152°F
T_6=142°F
60° BOILER FEEDWATER
HEAT RECOVERY
DECANTER
T_7=65°F
RECOVERED SOLVENT TO STORAGE
CONDENSED WATER VAPOR
T_4=250°F
92 PPM
120°F
WATER TO DRAIN

Figure 9. System configuration for refrigeration/condensation aersol removal control technology.

to obtain 95 percent overall control with 98 percent aerosol collection efficiency. The corresponding condenser outlet temperature from Figure 7 is 0°F.

Precooler Capacity--

The purified stack gas leaving the condenser/aerosol removal unit is essentially cold dry air. In the illustrated system this gas is used to precool the stack gas going to the condenser and provides the following cooling capacity:

$$\begin{aligned} H_{precooler} &= \dot{m}_a \; c_{p,a} \; (T_4 - T_3) \qquad (5) \\ &= 279 \times 0.24 \times (250 - 0) \\ &= 16{,}700 \text{ Btu/min} \end{aligned}$$

Condenser Inlet Gas Temperature--

The precooler capacity determined above will reduce the stack gas temperature from its initial value of 300°F. As a first approximation, assume that no condensation occurs in the precooler. The condenser inlet gas temperature is then

$$\begin{aligned} T_2 &= T_1 - \frac{H_{precooler}}{\dot{m}\, c_p} \qquad (6) \\ &= 300 - \frac{16{,}700}{(347)(0.29)} \\ &= 134°\text{F} \end{aligned}$$

If the precooler outlet gas temperature is 134°F, then some of the water vapor present in the exhaust gas would condense. Since condensation occurs in the precooler, then the outlet gas temperature is determined by:

$$T_2 = T_1 - \frac{(H_{precooler} - \dot{m}_{w,c}\, h_{fg,w})}{c_p\, \dot{m}} \qquad (7)$$

where $\dot{m}_{w,c}$ is the rate of water condensation and $h_{fg,w}$ is the heat of condensation. Since $\dot{m}_{w,c}$ and $h_{fg,w}$ are both functions of the outlet temperature, the above equation is solved by iteration.

For:

$$H_{precooler} = 16{,}700 \text{ Btu/min}$$

$$c_p = 0.29 \text{ Btu/(lb)(°F)}$$

$$\dot{m} = 347 \text{ lb/min}$$

the resulting precooler gas outlet temperature, T_2, is approximately 152°F and the rate of condensation in the precooler is less than 0.1 lb of water per minute.

Condenser Cooling Load--

With 95 percent VOC control, the rate of solvent condensation will vary between 0.4 and 14 lb/min for stack gas VOC concentrations between 200 and 8000 ppm. For all VOC concentrations of stack gas at 300°F and 6 percent relative humidity, however, the water vapor condensation rate will be approximately 60 lb/min. Since the heat of condensation for the solvent is an order of magnitude lower than that of water, solvent condensation is neglected in the following preliminary calculation of condenser cooling load:

$$\begin{aligned} H_{condenser} &= \dot{m}\, c_p\, (T_2 - T_3) + \dot{m}_w\, h_{fg,w} \qquad (8) \\ &= 347 \times 0.29 \times (152 - 0) + 63.7 \times 1050 \\ &= 82{,}200 \text{ Btu/min} \end{aligned}$$

Solvent Recirculation Rate--

The cooling load for the condenser is supplied by recirculating solvent in the system illustrated in Figure 9. This allows for easy recovery of the condensed solvent. For the specified solvent inlet and outlet temperatures, the solvent recirculation rate, $\dot{m}_{s,r}$, is:

$$\begin{aligned} \dot{m}_{s,r} &= \frac{H_{condenser}}{c_{p,s}\,(T_6 - T_5)} \qquad (9) \\ &= \frac{82{,}200}{(0.5)(142 + 10)} \\ &= 1080 \text{ lb/min} \end{aligned}$$

Refrigeration Load--

Again using the temperatures specified in Figure 9, the amount of refrigeration required for the system is determined by:

$$H_{refr} = \dot{m}_{s,r}\ c_{p,s}\ (T_7 - T_5) \qquad (10)$$

$$= 1080 \times 0.5 \times (65 + 10)$$

$$= 40{,}500 \text{ Btu/min}$$

This is expressed in typical refrigeration units by dividing H_{refr} by the conversion factor 200 (Btu/min)/ton. Thus,

$$H'_{refr} = 40{,}500/200$$

$$= 202 \text{ tons}$$

Waste Heat Recovery--

The preliminary system design includes a boiler feedwater/recirculating solvent heat exchanger to recover the thermal energy extracted from the stack gas. The energy savings of this unit are:

$$H_{wh} = [(\dot{m}_{s,r} + \dot{m}_{s,c})\ c_{p,s} + \dot{m}_{w,c}\ c_{p,w}](T_6 - T_7) \qquad (11)$$

$$= [(1080 + 5.3) \times 0.5 + 63.7 \times 1.0](142 - 65)$$

$$= 46{,}700 \text{ Btu/min}$$

where $\dot{m}_{s,c}$ and $\dot{m}_{w,c}$ are the mass flow rates of condensed solvent and water, respectively. The value of $\dot{m}_{s,c}$ is determined from the difference in exhaust gas VOC concentrations at the inlet and outlet of the condenser and ideal gas relations.

7.2.3 System Costs (3000 ppm case)

Capital Cost--

A textile plant raises its 10^6 gal/day water 9°F by using a similar water cooled condensing system. This equals 75.06×10^6 Btu/day. Hence, its condenser/waste heat recovery system cost of $87,100 can be used there.

1980 refrigeration system cost (18)	$239,800
Rest of equipment cost (see above)	87,100
Installation cost	63,200
Total installed equipment	390,100
Capital cost adjustment	115,000
Total	$505,100

Operating Cost--

Refrigeration--A refrigeration plant consumes approximately 1.5 kW per ton of refrigeration generated (18). The cost of power for the illustrated plant is therefore:

$$202 \text{ tons} \times 1.5 \, \frac{\text{kW}}{\text{ton}} \times 6000 \, \frac{\text{hr}}{\text{year}} \times \frac{\$0.055}{\text{kWhr}} = \$100{,}000/\text{year}$$

Annual operating costs for the illustrated system are thus:

Power cost	$100,000
Maintenance cost (10 percent of installed equipment)	39,000
Labor (0.5 man-years)	9,000
Total	$148,000/year

Operating Savings--

Value is recovered in operating the system by recovering the thermal energy of the recirculating solvent as it exits from the condenser and by utilizing the solvent recovered from the stack gas. The solvent may be reused in the textile operation or it may be used as fuel. The annual value of each of these savings is calculated below.

Waste Heat Recovery -

$$46{,}700 \, \frac{\text{Btu}}{\text{min}} \times \frac{\$2.9}{10^6 \text{ Btu}} \times \frac{60 \text{ min}}{\text{hr}} \times 6000 \, \frac{\text{hr}}{\text{year}} = \$48{,}800/\text{year}$$

Solvent for Reuse -

$$5.3 \, \frac{\text{lb}}{\text{min}} \times \frac{60 \text{ min}}{\text{hr}} \times 6000 \, \frac{\text{hr}}{\text{year}} \times \frac{\$0.18}{\text{lb}} = \$350{,}000/\text{year}$$

Solvent for Fuel -

$$5.3 \, \frac{\text{lb}}{\text{min}} \times \frac{60 \text{ min}}{\text{hr}} \times 6000 \, \frac{\text{hr}}{\text{year}} \times 19{,}930 \, \frac{\text{Btu}}{\text{lb}} \times \frac{\$2.9}{10^6 \text{ Btu}}$$

$$= \$110{,}000/\text{year}$$

7.2.4 System Variations with Concentration

Since neither the gas flow rate nor heat load vary widely, total capital cost, annual operating costs and waste heat recovery savings will remain *essentially* the same at all concentrations. The value of recovered product increases with increase in concentration. The resulting economic evaluation with operating costs and savings combined to give annual direct cost, is summarized under identification numbers I and II in Table 19.

TABLE 19. COSTS FOR REFRIGERATION/CONDENSATION AND AEROSOL REMOVAL SYSTEMS (BASED ON PRELIMINARY DESIGN)

Stack gas emissions concentration (ppm, volume)	Cost item	Identification Number I	Identification Number II
		Resource Recovery	
		Recovered solvent reused	Recovered solvent used as fuel
200	Total capital cost ($)	505,100	505,100
	Annual direct cost ($)	-75,200	-91,500
	Payback*	-	-
3000	Total capital cost ($)	505,100	505,100
	Annual direct cost ($)	+250,800	+10,800
	Payback*	2.0	47
8000	Total capital cost ($)	505,100	505,100
	Annual direct cost ($)	+820,800	+195,800
	Payback*	0.6	2.6
Notes	Negative annual direct cost indicates expenditure; positive annual direct cost indicates earned revenue	Stack gas not diluted in these cases.	

*Payback in years.

7.3 THERMAL INCINERATION SYSTEM

7.3.1 System Configuration

Many thermal incineration system configurations are possible (19). The configuration used in this study is shown in Figure 10.

7.3.2 Preliminary System Design (3000 ppm case)

Heat Required to Raise Stack Gas to Incineration Temperature--

As previously discussed in subsection 6.2, an upper limit for stack gas VOC concentration is imposed by the temperature limitations of materials upstream of a thermal incinerator. For a system with thermal energy recovery from the incinerator exhaust, as illustrated in Figure 10, the maximum allowable incinerator temperature, T_2, is approximately 1500°F. Thus, for nominal stack gas conditions of

$$\dot{m} = 347 \text{ lb/min}$$

$$T_1 = 300°\text{F}$$

$$c_p = 0.29 \text{ Btu/(lb)(°F)}$$

and

$$T_2 \leq 1500°\text{F}$$

the heat load of the incinerator, H_i, must be

$$\begin{aligned} H_i &\leq \dot{m}\, c_p\, (T_2 - T_1) \qquad (12) \\ &\leq 347 \times 0.29 \times (1500 - 300) \\ &\leq 1.21 \times 10^5 \text{ Btu/min} \end{aligned}$$

If heat recovery is not included in the system, the maximum allowable incinerator exhaust temperature increases to 2000°F. Thus, for $T_2 \leq 2000°\text{F}$

$$\begin{aligned} H_i &\leq 347 \times 0.29 \times (2000 - 300) \\ &\leq 1.71 \times 10^5 \text{ Btu/min} \end{aligned}$$

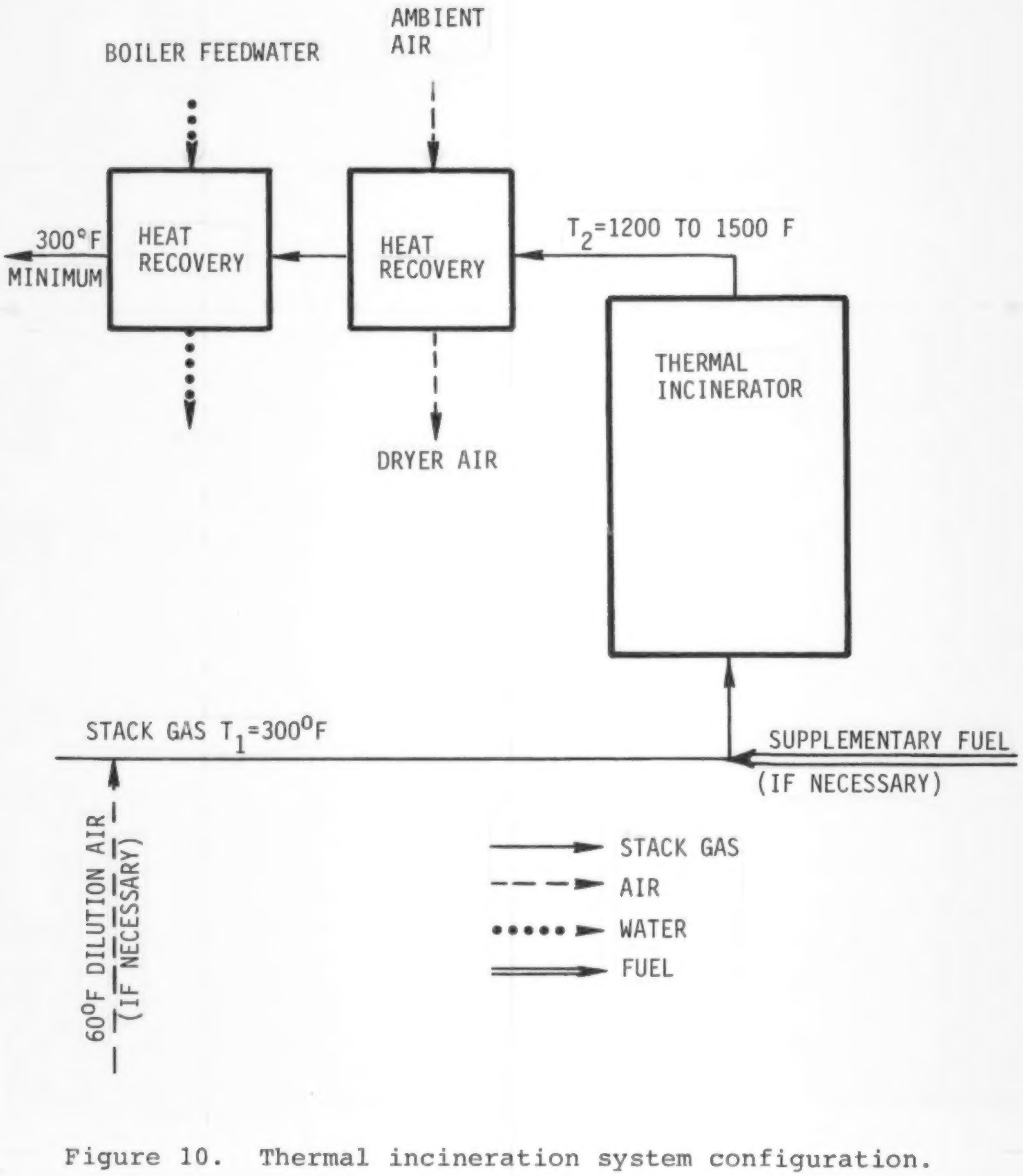
BOILER FEEDWATER
AMBIENT AIR
300°F MINIMUM
HEAT RECOVERY
HEAT RECOVERY
T2=1200 TO 1500 F
THERMAL INCINERATOR
DRYER AIR
STACK GAS T1=300°F
SUPPLEMENTARY FUEL (IF NECESSARY)
60°F DILUTION AIR (IF NECESSARY)
STACK GAS
AIR
WATER
FUEL

Figure 10. Thermal incineration system configuration.

A lower VOC concentration limit is also established by the autoignition temperature of 1200°F. Thus the same stack gas conditions and

$$T_2 \geq 1200°F$$

require that:

$$H_i \geq 347 \times 0.29 \times (1200 - 300)$$

$$\geq 9.06 \times 10^4 \text{ Btu/min}$$

Stack Gas Concentration Limit for Thermal Incineration--

These heat load boundaries are related to stack gas VOC concentration, C, by the following equation which determines the fuel value of the VOC's in the stack gas:

$$H = \frac{C}{10^6} \times Q \times \rho_s \times h_v \times \eta_i \tag{13}$$

where h_v is the solvent (Varsol) heating value in Btu/lb, and η_i is the incinerator efficiency. Thus,

$$C = \frac{H \times 10^6}{5000 \times 0.37 \times 19{,}930 \times 0.95} \tag{14}$$

and for thermal incineration with heat recovery,

$$2600 \leq C \leq 3500 \text{ ppm}$$

If heat recovery is not included in the system the maximum stack gas VOC concentration from Equation 14 is 4900 ppm. This exceeds 35 percent of the fuel lower explosive limit, or 3500 ppm for Varsol, which is a boundary practiced in industry. Thus, the concentration limits for thermal incineration with or without heat recovery are the same.

This shows that approximately 3000 ppm VOC concentration is required for thermal incineration for a stack gas inlet temperature of 300°F. Fuel must be added at concentrations below 2600 ppm.

From Equation 13 the fuel value of 200 ppm stack gas is only

$$H = \frac{200}{10^6} \times 5000 \times 0.37 \times 19{,}930 \times 0.95$$

$$\simeq 7000 \text{ Btu/min}$$

If natural gas with a heating value of 1000 Btu per standard cubic foot is used to supplement the solvent heat rate in the incinerator, then the natural gas flow rate must be

$$Q_f = \frac{90{,}600 - 7{,}000}{1{,}000}$$

$$= 84 \text{ scfm.}$$

Above 3500 ppm, to keep the incineration temperature below 1500°F with heat recovery and to remain below 35 percent of the lower explosion limit without heat recovery, the stack gases must be diluted.

The fuel value of VOC's in stack gases at 8000 ppm from Equation 13 is

$$H = \frac{8000}{10^6} \times 5000 \times 0.37 \times 19{,}930 \times 0.95$$

$$= 2.80 \times 10^5 \text{ Btu/min}$$

The amount of dilution air required to maintain $T_2 \leq 1500°\text{F}$ for a system with heat recovery is determined from

$$H = \dot{m}\, c_p\, (T_2 - T_1) + \dot{m}_{a,\text{dilution}}\, c_{p,a}\, (T_2 - T_a) \qquad (15)$$

Assuming the dilution air is dry ambient air at $T_a = 60°\text{F}$,

$$\dot{m}_{a,\text{ dilution}} = \frac{2.80 \times 10^5 - 355 \times 0.29 \times (1500 - 300)}{0.26 \times (1500 - 60)}$$

$$= 418 \text{ lb/min}$$

For a dry air density of 0.0764 lb/ft^3 under standard temperature and pressure (T = 60°F at one atmosphere),

$$Q_{a,\text{dilution}} = \frac{418}{0.0764} \simeq 5500 \text{ scfm}$$

for an approximate total adjusted flow rate of

$$Q' \simeq 10{,}500 \text{ scfm.}$$

By definition,

$$C = \frac{Q_s}{Q} \times 10^6. \qquad (16)$$

For the original conditions of C = 8000 ppm and Q = 5000 scfm, from Equation 16

$$Q_s = \frac{8000 \times 5000}{10^6} = 40 \text{ scfm}.$$

Given the same solvent flow rate and the adjusted total flow rate, the adjusted stack gas concentration is

$$C' = \frac{40}{10,500} \times 10^6 \simeq 3800 \text{ ppm}$$

which is greater than the concentration limit practiced by industry.

The dilution of 8000 ppm exhaust is therefore determined by the 3500 ppm maximum concentration limit rather than by temperature restrictions. Thus from Equation 16

$$Q' = \frac{40}{3500} \times 10^6 = 11,400 \text{ scfm}$$

and

$$Q_{a,dilution} = Q' - Q = 6400 \text{ scfm}$$

and from Equation 15

$$T_2 = \frac{H + \dot{m}\, c_p\, T_1 + \dot{m}_{a,dilution}\, c_{p,a}\, T_a}{\dot{m}\, c_p + \dot{m}_{a,dilution}\, c_{p,a}} = 1380°F$$

Incinerator Actual Airflow--

The capital cost of thermal incineration and heat recovery equipment is related to the size of the units. Incinerator size is a function of the residence time required to achieve the desired control efficiency at the incinerator temperature and actual volume flow rates. The gas volume changes in the incinerator resulting from the chemical reactions of the hydrocarbons and oxygen are small and therefore neglected here. To convert the adjusted inlet flow rate expressed in standard volumes, which imples 60°F and 14.7 psia, to actual volume flow rate, the ratio of absolute temperatures is used. That is,

$$Q^* = Q\left(\frac{T_2 + 460}{60 + 460}\right) \tag{17}$$

with Q = standard ft^3/min, scfm

Q^* = actual ft^3/min, acfm.

The actual volume of the incinerator exhaust which indicates the relative size of the emission control units and other pertinent thermal incinerator design conditions for each of the three concentration cases are reported in Table 20.

TABLE 20. INCINERATOR DESIGN CONDITIONS

Parameter	Stack gas concentration (ppm, volume) 200	 3,000	 8,000
Adjusted flow rate (scfm)	5,100	5,000	11,400
Incineration temperature (°F)	1,200	1,340	1,380
Heat load (Btu/min)	9.06×10^4	1.05×10^5	2.80×10^5
Fuel value (Btu/min)	0.70×10^4	1.05×10^5	2.80×10^5
Supplementary heat required (Btu/min)	8.36×10^4	-	-
Actual volume of incinerator outlet gases (acfm)	16,300	17,300	40,300

7.3.3 System Costs (3000 ppm case)

Capital Cost Without Heat Recovery--

1980 incinerator cost (9)	$36,700
Installation cost (9)	18,400
Total installed equipment	$55,100
Capital cost adjustment	16,200
Total	$71,300

Operating Cost Without Heat Recovery--

In order to overcome the system pressure losses, a fan or blower is used to deliver the stack gases to the incinerator at slightly more than atmospheric pressure. The power in kilowatts required for the blower is

$$P = 1.17 \times 10^{-4} Q \frac{\Delta p}{\eta_f} \tag{18}$$

where Δp is the system pressure drop in inches water gauge (in. WG) and η_f is the combined efficiency of the fan and fan motor. Assuming $\eta_f = 0.75$ and $\Delta p = 4$ in. WG for a thermal incineration system without heat recovery, the power required for the 3000 ppm stack gas case is

$$P = \frac{1.17 \times 10^{-4} \times 5000 \times 4}{0.75} = 3.1 \text{ kW}$$

The annual cost of power is then:

$$3.1 \text{ kW} \times 6000 \frac{\text{hr}}{\text{year}} \times \frac{\$0.055}{\text{kWhr}} \simeq \$1000/\text{year}$$

The annual operating costs without heat recovery are then:

Power cost	$ 1,000
Maintenance cost (10 percent of installed equipment)	5,500
Labor (0.5 man-years)	9,000
Total	$15,500/year

Capital Cost with Heat Recovery--

1980 incinerator cost (9)	$ 93,100
Installation cost (9)	41,900
Total installed equipment	$135,000
Capital cost adjustment	39,800
Total	$174,800

Operating Cost with Heat Recovery--

The addition of heat exchangers for energy recovery to the system will increase the labor charge and the system pressure losses. Assuming $\Delta p = 8$ in. WG for thermal incineration with heat recovery, the power required by the fan will be 6.2 kW from Equation 18 and the annual cost of power will be

$$6.2 \text{ kW} \times 6000 \frac{\text{hr}}{\text{year}} \times \frac{\$0.055}{\text{kWhr}} \simeq \$2000/\text{year}$$

so that

Power cost	$ 2,000
Maintenance cost (10 percent of installed equipment)	13,500
Labor (0.75 man-years)	13,500
Total	$29,000/year

Operating Savings with Heat Recovery--

The thermal energy of the incinerator exhaust may be recovered and used in place of burning fuel to provide heat for textile operations or power generation. The operating savings is thus the value of the fuel which would otherwise be required to supply the same thermal energy. With a minimum exhaust temperature of 300°F to prevent combustion product condensation, the maximum heat recoverable is

$$\begin{aligned} H &= \dot{m}\, C_p (T_2 - 300) \\ &= 347 \times 0.29 \times (1340 - 300) \\ &= 1.05 \times 10^5 \text{ Btu/min} \end{aligned}$$

and the annual operating savings due to recovered heat is

$$1.05 \times 10^5 \frac{\text{Btu}}{\text{min}} \times 60 \frac{\text{min}}{\text{hr}} \times 6000 \frac{\text{hr}}{\text{year}} \times \frac{\$2.9}{10^6 \text{ Btu}}$$

$$= \$109{,}600/\text{year}$$

7.3.4 System Variations with Concentration

System costs for the 200 and 8000 ppm cases are determined in a manner similar to the above evaluation. The cost varies primarily due to the differences in adjusted flow rates between the systems, which influences the capital costs and power costs. Also for the 200 ppm case the operating cost includes an annual fuel cost for the supplementary fuel required to achieve a 1200°F incinerator temperature.

The results of the preliminary cost analysis for thermal incineration without and with heat recovery are presented in Table 21 under identification numbers III and IV respectively.

7.4 FIXED BED CATALYTIC INCINERATION SYSTEM

7.4.1 System Configuration

The configuration assumed for evaluation of a fixed bed catalytic incineration system for hydrocarbon emission control is essentially the same as for the thermal incineration system illustrated in Figure 10. The differences between the two systems are noted below.

TABLE 21. COSTS FOR THERMAL INCINERATION SYSTEMS (BASED ON PRELIMINARY DESIGN)

Stack gas emissions concentration (ppm, volume)	Cost item	Identification number III	Identification number IV
		Resource recovery	
		No heat recovery	Heat recovery
200	Total capital cost ($)	73,200	165,600
	Annual direct cost ($)	-103,000	-21,100
	Payback*	-	-
3000	Total capital cost ($)	71,300	174,800
	Annual direct cost ($)	-15,500	+80,600
	Payback*	-	2.2
8000	Total capital cost ($)	106,000	324,100
	Annual direct cost ($)	-19,500	+249,100
	Payback*	-	1.3
Notes	Negative annual direct cost indicates expenditures; positive annual direct cost indicates earned revenue	Stack gas diluted to 3500 ppm, if higher	

*Payback in years.

7.4.2 Preliminary System Design (3000 ppm case)

The maximum allowable incinerator temperature is set by the catalyst material and not materials upstream of the incinerator. For this preliminary design the difference between the minimum and maximum allowable temperatures is assumed to be small and a single incinerator operating temperature, T_2 = 850°F is selected as the nominal design condition. The heat required to raise the nominal 347 lb/min of stack gas from 300°F to incineration temperature is therefore determined to be

$$H_i = 347 \times 0.29 \times (850 - 300)$$

$$= 5.53 \times 10^4 \text{ Btu/min}$$

from Equation 12, and the corresponding required stack gas concentration of VOC is

$$C = \frac{5.53 \times 10^4 \times 10^6}{5000 \times 0.37 \times 19{,}930 \times 0.95} = 1580 \text{ ppm}$$

from Equation 14.

Thus 200 ppm stack gas will require supplementary fuel while 3000 and 8000 ppm stack gases will require dilution prior to catalytic incineration.

The original fuel value of each stack gas concentration is the same regardless of VOC control used. Thus for 200 ppm the supplementary gas flow rate, using 1000 Btu/standard ft^3 natural gas, is

$$Q_f = \frac{55{,}300 - 7000}{1000} = 48 \text{ scfm.}$$

The amount of dilution air required for the 3000 and 8000 ppm cases is determined from Equation 15 to be

$$\dot{m}_{a,dilution} = \frac{1.05 \times 10^5 - 347 \times 0.29 \times (850 - 300)}{0.25 \times (850 - 60)}$$

$$= 250 \text{ lb/min for } C = 3000 \text{ ppm}$$

and

$$\dot{m}_{a,dilution} = \frac{2.80 \times 10^5 - 355 \times 0.29 \times (850 - 300)}{0.25 \times (850 - 60)}$$

$$= 1130 \text{ lb/min for } C = 8000 \text{ ppm.}$$

For $\rho_a = 0.0764$ lb/ft^3, this translates into dilution air flow rates of approximately 3300 and 14,800 scfm respectively.

Table 22 summarizes fixed bed catalytic incinerator design parameters for each of the stack gas VOC concentrations being evaluated.

TABLE 22. FIXED BED CATALYTIC INCINERATOR DESIGN CONDITIONS

	Stack gas concentration (ppm, volume)		
Parameter	200	3,000	8,000
Adjusted flow rate (scfm)	5,000	8,300	19,800
Incineration temperature (°F)	850	850	850
Heat load (Btu/min)	5.47×10^4	1.05×10^5	2.80×10^5
Fuel value (Btu/min)	0.70×10^4	1.05×10^5	2.80×10^5
Supplementary heat required (Btu/min)	4.77×10^4	-	-
Actual volume of incinerator outlet gases (acfm)	12,600	20,900	49,900

7.4.3 System Costs (3000 ppm case)

The costing procedure for a fixed bed catalytic incinerator is essentially the same as for a thermal incinerator, with the exception of the following cost assumptions:

- Maintenance cost is 15 percent of the installed equipment cost (9).
- Labor associated with the systems without and with heat recovery are 0.75 and 1.0 man-years respectively.

The costs are otherwise determined as in the previous thermal incinerator cost evaluations. The costs for the case of 3000 ppm VOC stack gas are summarized in the following tallies.

Capital Cost without Heat Recovery--

1980 incinerator cost (9)	$ 74,500
Installation cost (9)	37,200
Total installed equipment	$111,700
Capital cost adjustment	32,900
Total	$144,600

Operating Cost without Heat Recovery--

Power cost	$ 1,700
Maintenance cost (15 percent of installed equipment, includes catalyst replacement)	16,800
Labor (0.75 man-years)	13,500
Total	$32,000/year

Capital Cost with Heat Recovery--

1980 incinerator cost (9)	$128,100
Installation cost (9)	57,600
Total installed equipment	$185,700
Capital cost adjustment	54,700
Total	$240,400

Operating Cost with Heat Recovery--

Power cost	$ 3,500
Maintenance cost (15 percent of installed equipment, includes catalyst replacement)	27,900
Labor (1.0 man-years)	18,000
Total	$49,400/year

Operating Savings with Heat Recovery--

$$1.05 \times 10^5 \frac{\text{Btu}}{\text{min}} \times 60 \frac{\text{min}}{\text{hr}} \times 6000 \frac{\text{hr}}{\text{year}} \times \frac{\$2.9}{10^6 \text{Btu}}$$

$$= \$109{,}600/\text{year}$$

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7.4.4 System Variations with Concentration

As was the case for the analysis of the thermal incineration system, the increase in capital cost with VOC concentration is primarily due to the increase in adjusted flow rate and therefore equipment size. Also the increase in thermal savings with concentration causes an increase in direct costs (earned revenue) with concentration for the system with heat recovery. Operating for the 200 ppm concentration includes the cost of fuel required to achieve combustion temperature in both cases.

The results of the preliminary cost analysis of the fixed bed catalytic incineration system without and with heat recovery are presented in Table 23 under identification numbers V and VI, respectively.

7.5 FLUIDIZED BED CATALYTIC INCINERATION SYSTEM

7.5.1 System Configuration

The system configuration for fluidized bed catalytic incineration is illustrated in Figure 11.

7.5.2 Preliminary System Design (3000 ppm case)

The design assumptions which distinguish this system from either a fixed bed or thermal incineration system are that:

- Stack gases preheated to 800°F must contact the catalyst before combustion of VOC takes place.
- In-bed heat transfer tubes can be used to remove excess heat value of VOCs.
- The two assumptions above permit operations at any concentration below LEL. With thermal or fixed bed catalytic incineration, the operations must be below 35 percent LEL.
- Gas flow rate through the bed must meet fluidizing velocity and residence time requirements (20).

The stack gas content must be adjusted by the addition of dilution air to prevent overheating the incinerator when VOC concentrations are high and there is no heat recovery in the system. For example, the dilution air requirement for 5000 scfm of 3000 ppm VOC stack gas is determined from Equation 15 to be

TABLE 23. COSTS FOR FIXED BED CATALYTIC INCINERATION SYSTEMS (BASED ON PRELIMINARY DESIGN)

Stack gas emissions concentration (ppm, volume)	Cost item	Identification number V Resource recovery No heat recovery	Identification number VI Resource recovery Heat recovery
200	Total capital cost ($)	105,800	185,700
	Annual direct cost ($)	-76,600	-34,300
	Payback*	-	-
3000	Total capital cost ($)	144,600	240,400
	Annual direct cost ($)	-32,000	+60,200
	Payback*	-	4.0
8000	Total capital cost ($)	249,400	469,300
	Annual direct cost ($)	-46,600	+211,500
	Payback*	-	2.2
Notes	Negative annual direct cost indicates expenditure; positive annual direct cost indicates earned revenue	Stack gas diluted to 1580 ppm, if higher	

*Payback in years.

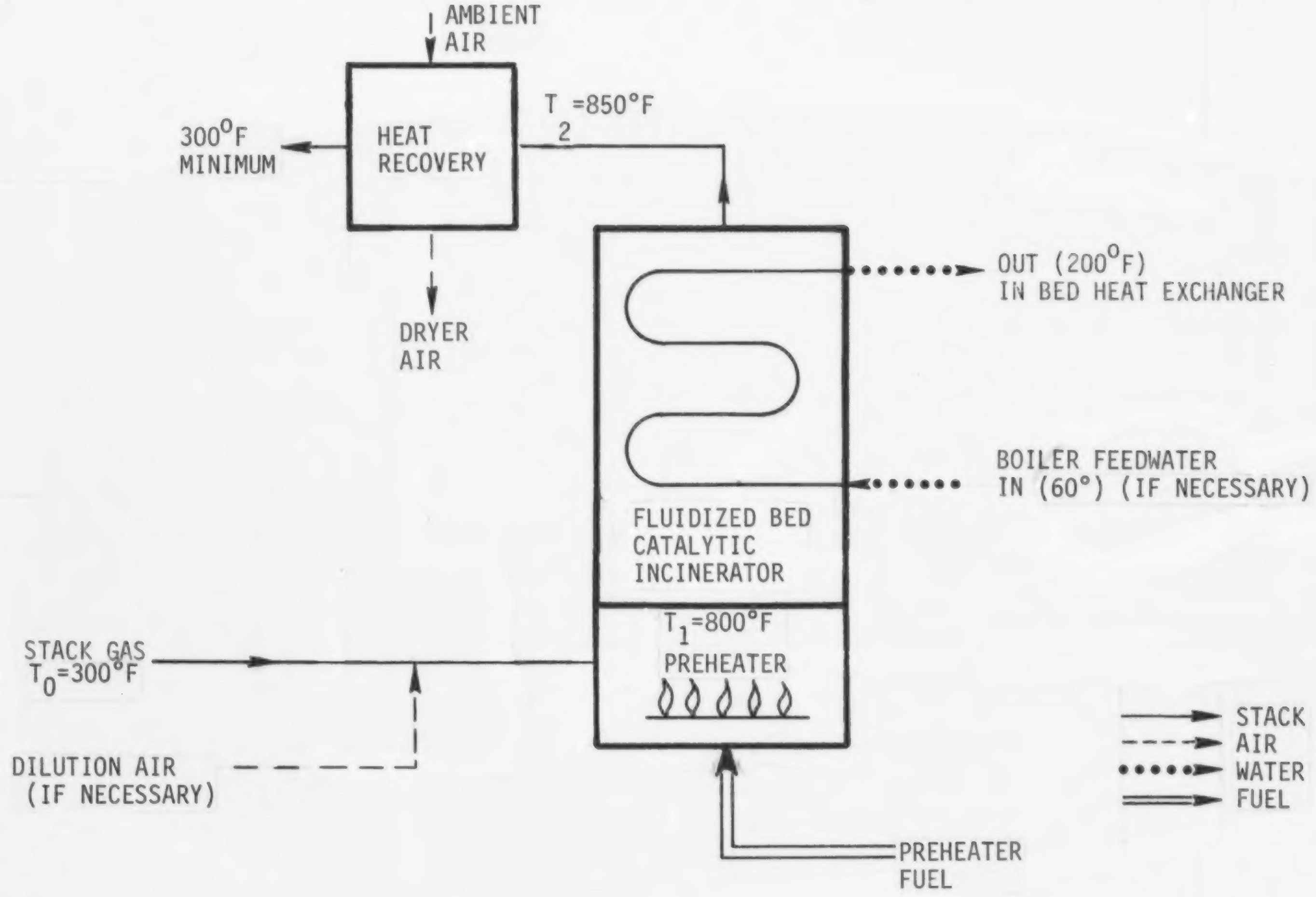
AMBIENT AIR
300°F MINIMUM
HEAT RECOVERY
T_2 =850°F
DRYER AIR
OUT (200°F) IN BED HEAT EXCHANGER
BOILER FEEDWATER IN (60°) (IF NECESSARY)
FLUIDIZED BED CATALYTIC INCINERATOR
T_1=800°F
PREHEATER
STACK GAS T_0=300°F
DILUTION AIR (IF NECESSARY)
PREHEATER FUEL
STACK
AIR
WATER
FUEL

Figure 11. Fluidized bed catalytic incinerator system configuration.

$$\dot{m}_{a,dilution} = \frac{1.05 \times 10^5 - 347 \times 0.29 \times (850 - 800)}{0.26 \times (850 - 800)}$$

$$= 7690 \text{ lb/min}$$

At 0.0764 lb of dry air per standard cubic foot this converts to 101,000 scfm which is more than 20 times the original flow rate. Fluidized bed incineration is obviously impractical without in-bed heat exchangers for catalyst cooling except for very low concentrations.

When in-bed heat recovery is included in the system, no dilution is required as long as VOC concentrations are below LEL. The following analysis is therefore restricted to fluidized bed incineration systems with heat recovery including in-bed heat exchangers for catalyst cooling.

In-Bed Heat Exchanger--

Since a preheater raises the stack gas from 300° to 800°F before it reaches the incinerator, the VOCs need only provide the heat value necessary to raise the gas temperature from 800° to 850°F. Thus the heat load of the catalytic incinerator is only H_i = 5000 Btu/min from Equation 12.

The heat available from 95 percent efficient combustion of 3000 ppm VOC exhaust gas is $H = 1.05 \times 10^5$ Btu/min as previously determined. The in-bed heat exchanger must therefore remove $H - H_i = 1.00 \times 10^5$ Btu/min from the fluidized bed to prevent overheating. Assuming boiler feedwater is being raised from 60° to 200°F in this process as is illustrated in Figure 11 and the catalyst temperature is constant at 850°F, the in-bed heat exchanger area is

$$A = \frac{(H - H_i) \times 60}{U \times \Delta T_m} \qquad (19)$$

$$= \frac{1.0 \times 10^5 \times 60}{35 \times 720} = 240 \text{ ft}^2$$

where U = heat transfer coefficient, Btu/(hr)(ft^2)(°F)

ΔT_m = mean temperature difference across the heat exchanger, °F

Further, assuming a tube side velocity, u_t, of approximately 500 ft/min, an appropriate heat exchanger tube diameter, d, and the catalyst bed volume occupied by the tubes, V_t, can be determined as follows:

$$\dot{m}_w = \frac{H - H_i}{c_{p,w}\ \Delta T_w} \tag{20}$$

$$= \frac{1.00 \times 10^5}{1.0 \times (200 - 60)} = 714 \text{ lb/min}$$

$$d = \sqrt{\frac{4(144)}{\pi}\ \frac{\dot{m}_w}{\rho_w\ u_t}} \tag{21}$$

$$= \sqrt{\frac{4(144)(714)}{\pi(61.5)(500)}} \simeq 2 \text{ in.}$$

$$V_t = \frac{12A}{\pi d} \times \frac{\pi d^2}{4(144)} = \frac{Ad}{48} \tag{22}$$

$$= \frac{240 \times 2}{48} = 10 \text{ ft}^3$$

Sizing of Fluidized Bed Incinerator--

Since dilution of the stack gases is not required when an in-bed heat recovery unit is used to control the operating temperature of the fluidized bed, the actual flow rate of the stack gases at any of the concentrations being considered is

$$Q^* = 5000\ \frac{(850 + 460)}{(60 + 460)} = 12{,}600 \text{ acfm}$$

according to Equation 17.

A procedure (20) for preliminary design of a fluidized bed catalytic incinerator with in-bed heat exchanger is illustrated below for the 3000 ppm VOC concentration case.

For a fluidizing velocity, $u_f = 150$ ft/min, and residence time, $t_r = 0.5$ sec:

- Fluidized catalyst bed height, $X_b = u_f t_r/60 = 1.25$ ft
- Incinerator cross-sectional area, $A_i = Q^*/u_f + V_t/X_b = 92 \text{ ft}^2$
- Incinerator diameter, $D_i = \sqrt{4\ A_i/\pi} = 10.8$ ft
- Disengagement height, $X_d = 3\ X_b = 3.75$ ft

- Gas inlet plenum height, $X_p = D_i/3 = 3.6$ ft
- Total chamber height, $X = X_b + X_d + X_p = 8.6$ ft
- Weight of chamber in 12 gauge (thickness, $t = 0.109$ in.) carbon steel (density, $\rho = 0.284$ lb/in.^{3}), including a distributor:

$$W = 144\ \rho t \left(\pi DX + \frac{3\pi D^2}{4}\right) = 2530 \text{ lb}$$

- Pressure drop through fluidized bed, with
  - Catalyst specific gravity, $\gamma = 2.7$
  - Catalyst voidage (fraction of unit stationary bed volume occupied by catalyst), $\alpha = 0.5$
  - Expanded bed voidage (fraction of unit fluidized bed volume occupied by catalyst), $\beta = 0.48$

  is determined by, $\Delta p_b = 12\ \alpha\beta\gamma X_b = 9.7$ in. WG
- Distributor pressure drop, $\Delta p_d = \Delta p_b/2 = 4.9$ in. WG
- Particulate collector pressure drop, $\Delta p_c = 4.0$ in. WG
- Preheater pressure drop, $\Delta p_p = 4.0$ in. WG
- Heat recovery exchanger pressure drop, $\Delta p_E = 4.0$ in. WG
- Total system pressure drop, $\Delta p = \Delta p_b + \Delta p_d + \Delta p_c + \Delta p_p + \Delta P_E = 26.6$ in. WG
- Catalyst inventory in-bed, $(\pi D^2/4)\ \alpha\beta\gamma\rho_w\ X_b = 4630$ lb.

7.5.3 System Costs (3000 ppm case)

Capital Cost with Heat Recovery--

Fabricated chamber cost (at $2.5/lb)	$ 6,300
Collector cost (at $2.5/acfm)	31,500
Basic equipment cost	$37,800

Allowing 60 percent of the basic equipment cost for ancillaries such as ducts, blower, instrumentation, catalyst, etc.:

Ancillaries	$ 22,700
In-bed heat exchanger ($40/ft^2)	4,100
Exhaust heat recovery exchanger	44,400
Total equipment cost	$109,000
Installation cost	64,600
Total installed equipment	$173,600
Capital cost adjustment	51,200
Total	$224,800

Operating Cost with Heat Recovery--

The annual cost of the fuel required to preheat the stack gas to 800°F is

$$347\ \frac{\text{lb}}{\text{min}} \times 0.29\ \frac{\text{Btu}}{(\text{lb})(^\circ\text{F})} \times (800 - 300)^\circ\text{F} \times 60\ \frac{\text{min}}{\text{hr}}$$

$$\times\ 6000\ \frac{\text{hr}}{\text{year}} \times \frac{\$2.9}{10^6\ \text{Btu}} = \$52,500/\text{year}$$

Thus,

Fuel cost	$52,500
Power cost	6,800
Maintenance cost (15 percent of installed equipment, includes catalyst replacement)	26,000
Labor (0.75 manyears)	13,500
Total	$98,800/year

Operating Savings with Heat Recovery--

The in-bed heat exchanger and incinerator exhaust stack heat recovery exchanger will produce an operating savings equivalent to the heating value of the preheat fuel and the stack gas VOCs.

$$(50,300 + 105,100)\ \frac{\text{Btu}}{\text{min}} \times 60\ \frac{\text{min}}{\text{hr}} \times 6000\ \frac{\text{hr}}{\text{year}}$$

$$\times\ \frac{\$2.9}{10^6\ \text{Btu}} = \$162,100/\text{year}$$

7.5.4 System Variations with Concentration

Since the equipment size is based on fluidizing velocity, the only additional equipment cost for higher concentration control systems is the in-bed heat exchanger. The economic evaluation of the fluidized bed catalytic incinerator is summarized under identification number VII in Table 24.

Ancillaries	$ 22,700
In-bed heat exchanger ($40/ft^2)	4,100
Exhaust heat recovery exchanger	44,400
Total equipment cost	$109,000
Installation cost	64,600
Total installed equipment	$173,600
Capital cost adjustment	51,200
Total	$224,800

Operating Cost with Heat Recovery--

The annual cost of the fuel required to preheat the stack gas to 800°F is

$$347\ \frac{\text{lb}}{\text{min}} \times 0.29\ \frac{\text{Btu}}{(\text{lb})(^\circ\text{F})} \times (800 - 300)^\circ\text{F} \times 60\ \frac{\text{min}}{\text{hr}}$$

$$\times\ 6000\ \frac{\text{hr}}{\text{year}} \times \frac{\$2.9}{10^6\ \text{Btu}} = \$52{,}500/\text{year}$$

Thus,

Fuel cost	$52,500
Power cost	6,800
Maintenance cost (15 percent of installed equipment, includes catalyst replacement)	26,000
Labor (0.75 manyears)	13,500
Total	$98,800/year

Operating Savings with Heat Recovery--

The in-bed heat exchanger and incinerator exhaust stack heat recovery exchanger will produce an operating savings equivalent to the heating value of the preheat fuel and the stack gas VOCs.

$$(50{,}300 + 105{,}100)\ \frac{\text{Btu}}{\text{min}} \times 60\ \frac{\text{min}}{\text{hr}} \times 6000\ \frac{\text{hr}}{\text{year}}$$

$$\times\ \frac{\$2.9}{10^6\ \text{Btu}} = \$162{,}100/\text{year}$$

7.5.4 System Variations with Concentration

Since the equipment size is based on fluidizing velocity, the only additional equipment cost for higher concentration control systems is the in-bed heat exchanger. The economic evaluation of the fluidized bed catalytic incinerator is summarized under identification number VII in Table 24.

TABLE 24. COSTS FOR FLUIDIZED BED CATALYTIC INCINERATION SYSTEMS (BASED ON PRELIMINARY DESIGN)

Stack gas emissions concentration (ppm, volume)	Cost item	Identification number VII Resource recovery With heat recovery
200	Total capital cost ($)	214,300
	Annual direct cost ($)	-37,800
	Payback*	-
3000	Total capital cost ($)	224,800
	Annual direct cost ($)	+63,300
	Payback*	3.6
8000	Total capital cost ($)	256,500
	Annual direct cost ($)	+242,500
	Payback*	1.1
Notes	Negative annual direct cost indicates expenditure; positive annual direct cost indicates earned revenue	Stack gas not diluted. In-bed heat exchanger provided for temperature control.

*Payback in years.

7.6 FIXED BED CARBON ADSORPTION

In describing adsorption as an applicable control technology, it was pointed out that the adsorption capacity of carbon is a function of VOC type, concentration and temperature. In general, longer chain, higher boiling point compounds, higher concentrations and lower temperatures increase adsorption capacity also referred to as adsorption efficiency (expressed in pounds VOC adsorbed/100 lb of carbon).

Figure 12 shows carbon adsorption design efficiencies, η_{ad}, for 95 percent control of several hydrocarbons at 100°F (9). As can be seen, the greater the number of carbon atoms in the compound, the higher the efficiency. An efficiency of 12.5 lb/100 lb has been extrapolated for a C_{10} hydrocarbon. These efficiencies are generally satisfactory for estimating purposes for concentrations of 25 ppm and greater. To illustrate the variation of system economics with concentration however, these values are assumed to apply to 3000 ppm exhaust and design efficiencies are assumed to vary with concentration as toluene equilibrium efficiencies vary with concentration (Figure 8). The equilibrium concentration for toluene at 100°F and 3000 ppm is 31 lb/100 lb. Thus the ratio of design efficiency for 95 percent control of C_{10} hydrocarbons to the equilibrium concentration of toluene is 0.40. This ratio is assumed to be constant for all temperatures and concentrations. The resulting design isotherms are illustrated in Figure 13 and are used in the following analyses.

In multicomponent adsorption, the higher boiling compounds tend to displace already adsorbed lower boiling ones, resulting in a complex mechanism which can only be established by actual tests. This design does not account for this complex adsorption mechanism.

The presence of moisture also affects the overall efficiency. However, adsorbers are in operation at relative humidities of up to 50 percent at 100°F. Accordingly, the effect of moisture has also been omitted in this study.

7.6.1 System Configuration

The system configuration is shown in Figure 14.

7.6.2 Preliminary System Design (3000 ppm case)

Emissions control by fixed bed adsorption is actually a three-step process which consists of adsorption, regeneration and fixed bed cooling processes in sequence.

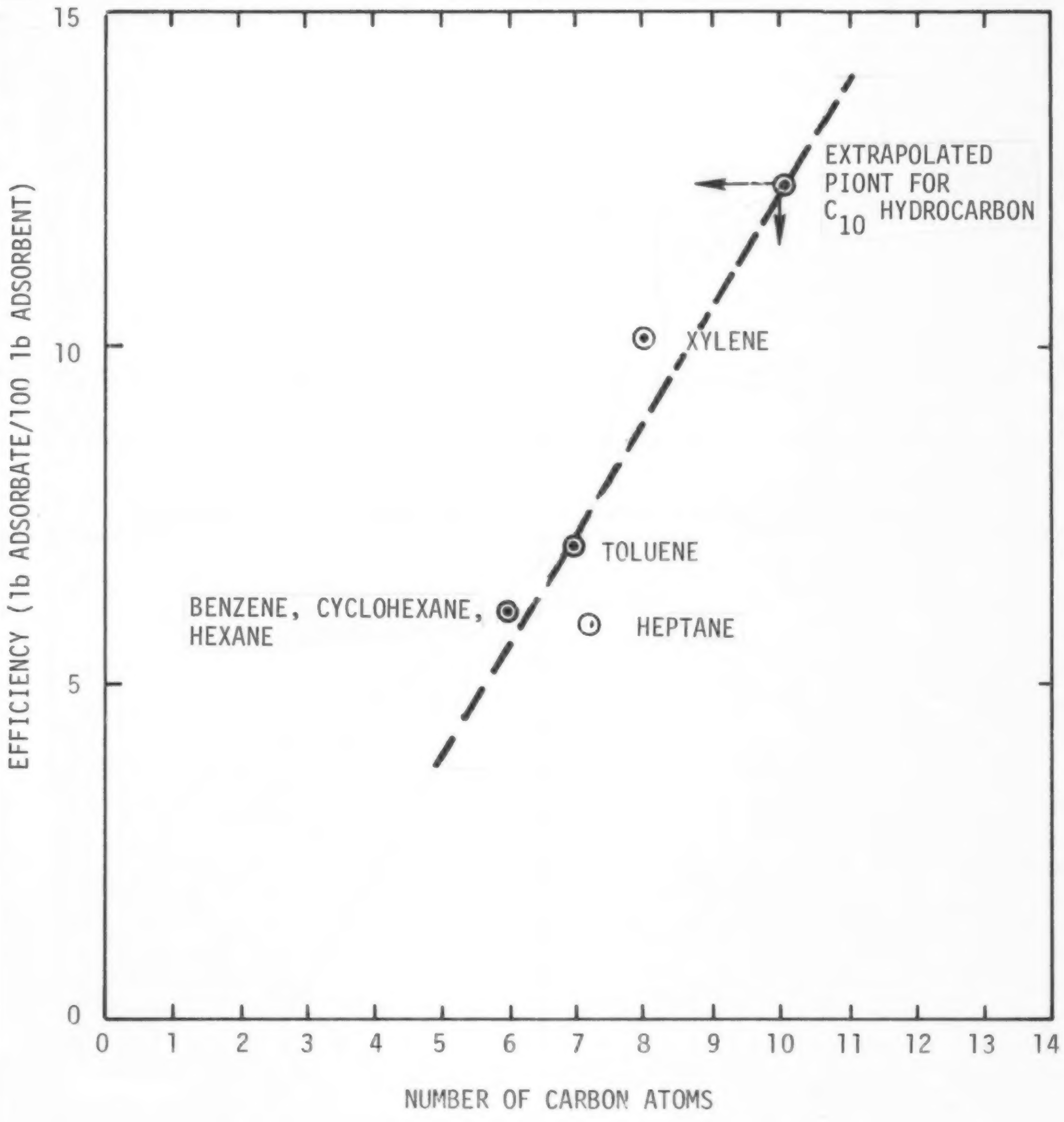

Figure 12. Variation of 95 percent control design adsorption efficiency with number of carbon atoms in hydrocarbon.

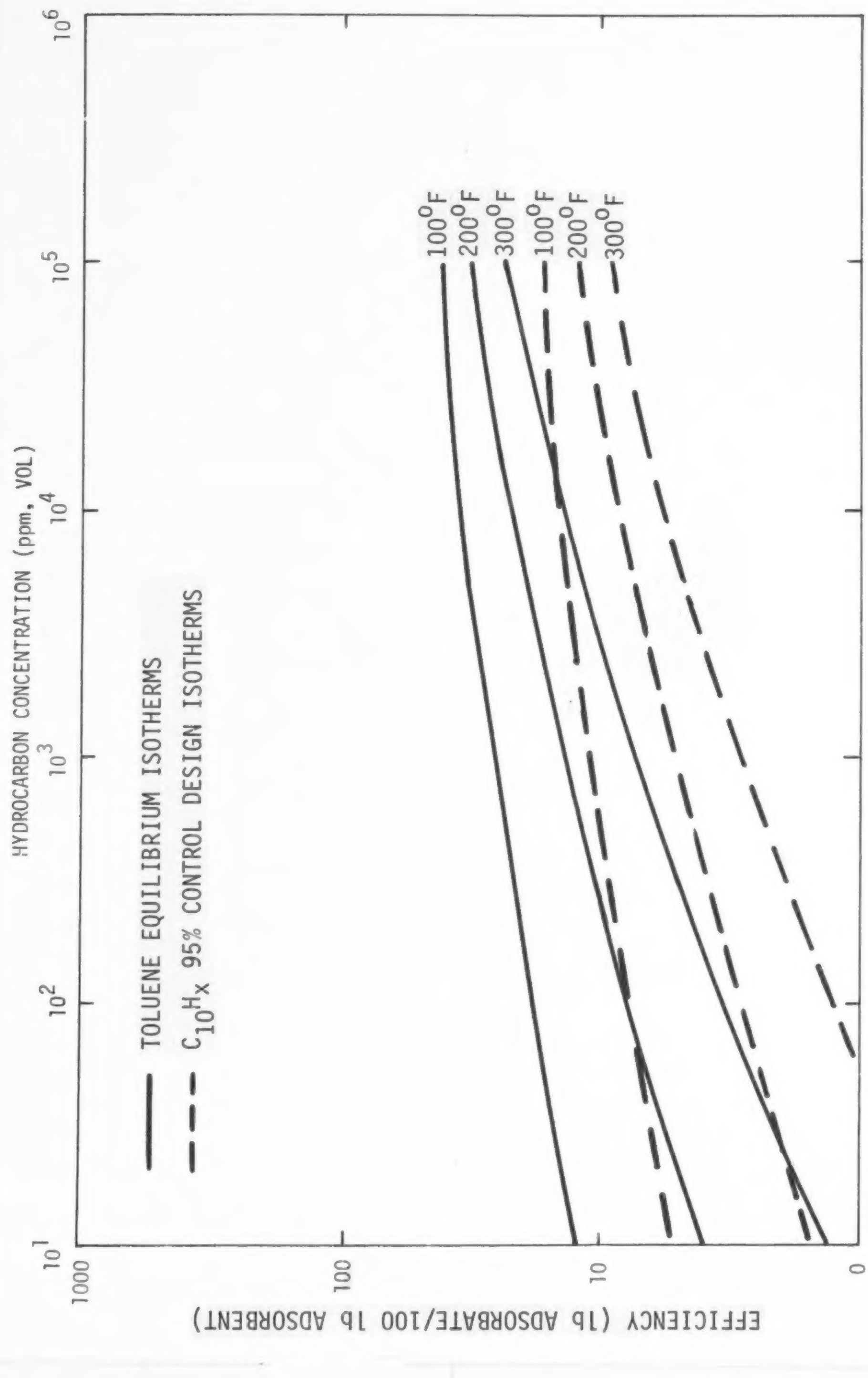

Figure 13. A comparison of $C_{10}H_x$ design and toluene equilibrium adsorption efficiencies for fixed bed adsorption.

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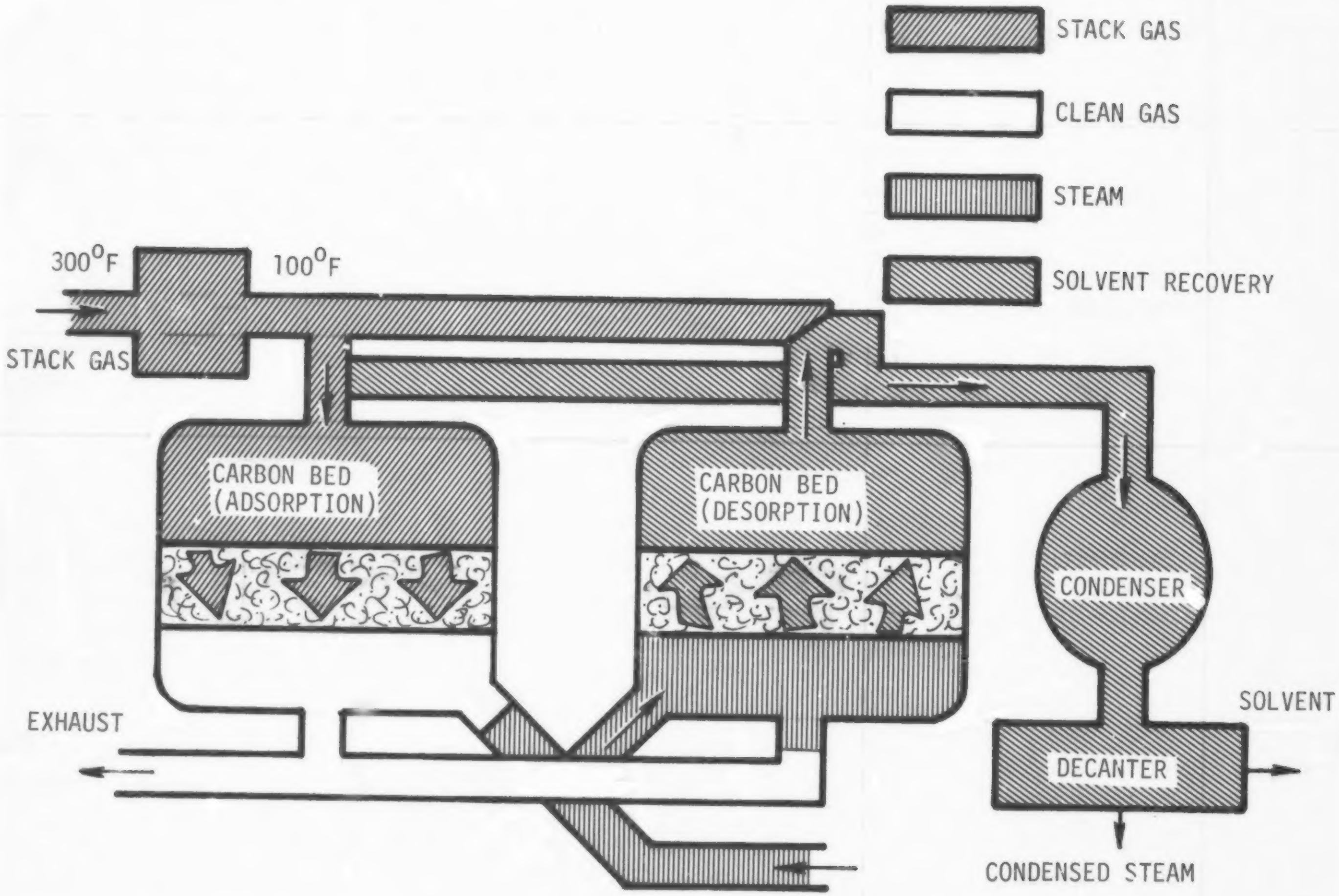
STACK GAS
CLEAN GAS
STEAM
SOLVENT RECOVERY
300°F
100°F
STACK GAS
CARBON BED (ADSORPTION)
CARBON BED (DESORPTION)
CONDENSER
SOLVENT
DECANTER
CONDENSED STEAM
EXHAUST

Figure 14. Fixed bed adsorber system configuration.

Adsorption--

Fixed bed adsorption systems are typically sized (9) to a specific gas flow rate of 200 (ft^3/min)/100 lb carbon at 100°F (186 scfm/100 lb carbon). Thus for 5000 scfm stack gas flow rate, the quantity of carbon required is

$$m_c = \frac{5000 \text{ scfm}}{186 \text{ scfm}/100 \text{ lb carbon}} \simeq 2690 \text{ lb.}$$

Ninety-five percent effective emission control implies that the solvent adsorption rate must be 0.95 $\dot{m}_s$. The appropriate design adsorption efficiency, assuming the stack gas is precooled to 100°F to improve efficiency, was previously determined to be 12.5 lb solvent/100 lb carbon. The above information permits the determination of the cycle time for adsorption, t_{ad}, from the following:

$$t_{ad} = \frac{m_c \, \eta_{ad}}{0.95 \, \dot{m}_s} \tag{23}$$

$$= \frac{2690 \times 12.5/100}{0.95 \times 5.5} \simeq 64 \text{ min.}$$

Regeneration--

Typically low pressure saturated steam is passed through the fixed bed in the direction opposite that of the stack gas flow for regeneration. Steam consumption, SC (lb steam/lb solvent), varies with solvent adsorbed and elapsed time. Based on data presented in (17) for fixed bed Naphtha adsorption, a steam consumption rate of 5.2 lb steam/lb solvent has been assumed for Varsol since the two substances have similar molecular weights, boiling points and latent heats.

The steam rate, $\dot{m}_{st}$, required for regeneration, given

- Saturated steam at 10 psig (240°F)
- Steam density, $\rho_{st} = 0.061 \text{ lb/ft}^3$
- A specific steam flow rate (9), $Q_{st} = 30$ (ft^3/min)/100 lb carbon

is equal to:

$$\dot{m}_{st} = Q_{st} \, \rho_{st} \, m_c \tag{24}$$

$$= \left(\frac{30}{100}\right) \times 0.061 \times 2690 \simeq 49 \text{ lb/min}$$

The cycle time for regeneration can also be determined:

$$t_{regen} = \frac{\eta_{ad}\ SC}{Q_{st}\ \rho_{st}} \qquad (25)$$

$$= \frac{(12.5/100) \times 5.2}{(30/100) \times 0.061} = 36 \text{ min}$$

Fixed Bed Cooling--

The thermal energy transfer to the carbon bed during regeneration must be removed before adsorption can again occur. For the following carbon bed characteristics:

- Carbon specific heat, $c_{p,c} = 0.225$ Btu/(lb)(°F)
- Bed temperature before cooling = 240°F
- Bed temperature after cooling = 100°F

and neglecting the heat of adsorption, the carbon bed cooling load is

$$H = \frac{m_c\ c_{p,c}\ \Delta T_c}{t_c} \qquad (26)$$

$$= \frac{2690 \times 0.225 \times 140}{t_c} = \frac{84{,}700}{t_c} \text{ Btu/min}$$

where t_c is the cycle time allotted for fixed bed cooling. Within limits the cooling system may be designed so that the time required for cooling complements the time required for the other cycle processes in order to minimize system costs.

For continuous emission control or adsorption, two or more fixed beds are required. With multiple units regeneration and cooling can take place in one or more beds while adsorption occurs in another. The optimum number of beds is indicated by the ratio of complete cycle time to adsorption process time. For the above case, the number of carbon beds required, N, is approximately

$$N = \frac{t_{ad} + t_{regen} + t_c}{t_{ad}} \qquad (27)$$

$$= \frac{64 + 36 + t_c}{64}$$

Thus t_c must be less than or equal to 28 min for continuous emission control with a two bed adsorption system. The resulting carbon bed cooling load is then approximately 3000 Btu/min which is easily achieved by circulating cooling air or nitrogen through the bed or by water-cooled coils in the bed.

The total carbon inventory for a two bed adsorption system is therefore 2 × 2690 = 5380 lb of carbon, and the replacement rate (assuming 10 percent annual loss) is 538 lb/year.

7.6.3 System Costs (3000 ppm case)

Capital Cost--

1980 fixed bed adsorption system cost (9)	$ 72,100
Installation cost (9)	36,000
Total installed equipment	$108,100
Capital cost adjustment	31,900
Total	$140,000

Operating Cost--

Operation of a fixed bed adsorption system requires: power for fans, blowers, pumps and automatic controls; cooling water for stack gas precooling, carbon bed cooling and condenser operation; and steam for carbon bed desorption.

The steam requirement has already been determined to be 5.2 lb steam/lb solvent adsorbed. With 95 percent control the annual steam rate is

$$0.95 \times 5.5 \ \frac{\text{lb solvent}}{\text{min}} \times 5.2 \ \frac{\text{lb steam}}{\text{lb solvent}} \times 60 \ \frac{\text{min}}{\text{hr}}$$

$$\times \ 6000 \ \frac{\text{hr}}{\text{year}} = 9.78 \times 10^6 \ \frac{\text{lb steam}}{\text{year}}$$

The requirements for power and cooling water have been determined in proportion to the requirements for the Naphtha adsorption system of (17). Power is assumed to be proportional to gas flow rate and cooling water to solvent recovered. The following annual utility usage rates result:

- Power - 21.1 kW × 6000 hr/year = 126,600 kWhr/year
- Cooling water - 43.8 tons/hr × 6000 hr/year = 262,800 tons/year.

Thus,

Steam cost ($4/1000 lb)	$39,100
Power cost	7,000
Cooling water cost ($0.035/ton (17))	9,200
Labor cost (0.5 man-years)	9,000
Maintenance cost	5,800
Carbon replacement cost	6,500
Total annual operating cost	$76,600/year

Operating Savings with Solvent Reuse--
The solvent reuse value is as follows:

- Solvent recovered:

$$5.5\ \frac{\text{lb solvent}}{\text{min}} \times 60\ \frac{\text{min}}{\text{hr}} \times 0.95 = 314\ \text{lb/hr}$$

- Reuse value: $0.18/lb

Recovered solvent reuse value is

$$314\ \frac{\text{lb}}{\text{hr}} \times \frac{\$0.18}{\text{lb}} \times 6000\ \frac{\text{hr}}{\text{year}} = \$339{,}100/\text{year}$$

Operating Savings with Solvent as Fuel--
The solvent fuel value is as follows:

- Solvent recovered:

$$5.5\ \frac{\text{lb solvent}}{\text{min}} \times 60\ \frac{\text{min}}{\text{hr}} \times 0.95 = 314\ \text{lb/hr}$$

- Fuel value: $0.058/lb

Recovered solvent fuel value is

$$314\ \frac{\text{lb}}{\text{hr}} \times \frac{\$0.058}{\text{lb}} \times 6000\ \frac{\text{hr}}{\text{year}} = \$109{,}300/\text{year}$$

7.6.4 System Variations with Concentration

To meet residence time requirements and optimize pressure drop through the bed, industry recommends a gas flow rate of 186 scfm/100 lb of carbon (9). For the 200 ppm concentration case to meet the gas flow rate condition the equipment size will remain the same as that for the 3000 ppm case. However, the system can adsorb on a single bed up to 595 min before regeneration

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is required versus 65 min for 3000 ppm stack gas. Operating costs for the 200 ppm case are reduced due to lower steam and cooling water requirements as are operating savings due to less recovered solvent.

Fixed beds do not lend themselves to economical indirect cooling, because of the low heat transfer rates. To avoid dangers from overheating due to the heat of adsorption, which could result in fires or explosion, higher concentration stack gases must be diluted. In this study, 3500 ppm (35 percent LEL) has been used as the upper limit, and the 8000 ppm case has been treated as a higher volume 3500 ppm case.

The resulting economic evaluation is summarized under identification numbers VIII and IX in Table 25.

7.7 FLUIDIZED BED ADSORPTION

Reference (17) reports a solvent recovery rate of 100 kg/hr with a carbon circulation rate of 600 kg/hr. This implies a design adsorption efficiency of 16.7 lb toluene/100 lb carbon for the 1680 ppm, 116°F inlet gas conditions. This is 65 percent of the toluene equilibrium adsorption efficiency for the same concentration and temperature (that is, 25.7 lb toluene/100 lb carbon from Figure 13). Assuming the same ratio of design to equilibrium efficiency, the toluene design point for 3000 ppm at 100°F is 20.1 lb toluene/100 lb carbon, which is 2.87 times the fixed bed toluene design efficiency of 7 lb toluene/100 lb carbon (9) illustrated in Figure 12. This improvement in design adsorption efficiency for fluidized versus fixed beds is assumed to apply to C_{10} hydrocarbon adsorption as well. The resulting fluidized bed design efficiencies used in the following analyses are presented in Figure 15 versus concentration for 100°F.

7.7.1 System Configuration

The fluidized bed adsorber system configuration is shown in Figure 16.

7.7.2 Preliminary Design Procedure (3000 ppm case)

The design efficiency for fluidized bed adsorption of 3000 ppm stack gas precooled to 100°F from Figure 15 is 35.9 lb solvent/100 lb carbon. Thus for an emissions control efficiency of 95 percent the carbon circulation rate is:

TABLE 25. COSTS FOR FIXED BED CARBON ADSORPTION SYSTEMS (BASED ON PRELIMINARY DESIGN)

Stack gas emissions concentration (ppm, volume)	Cost item	Identification number VIII	Identification number IX
		Resource recovery	
		Recovered solvent reused	Recovered solvent used as fuel
200	Total capital cost ($)	140,000	140,000
	Annual direct cost ($)	-7,200	-23,900
	Payback*	-	-
3000	Total capital cost ($)	140,000	140,000
	Annual direct cost ($)	+262,500	+32,700
	Payback*	0.5	4.3
8000	Total capital cost ($)	259,600	259,600
	Annual direct cost ($)	+734,300	+121,000
	Payback*	0.4	2.1
Notes	Negative annual direct cost indicates expenditure; positive annual direct cost indicates earned revenue	Stack gas diluted to 3500 ppm, if higher	

*Payback in years.

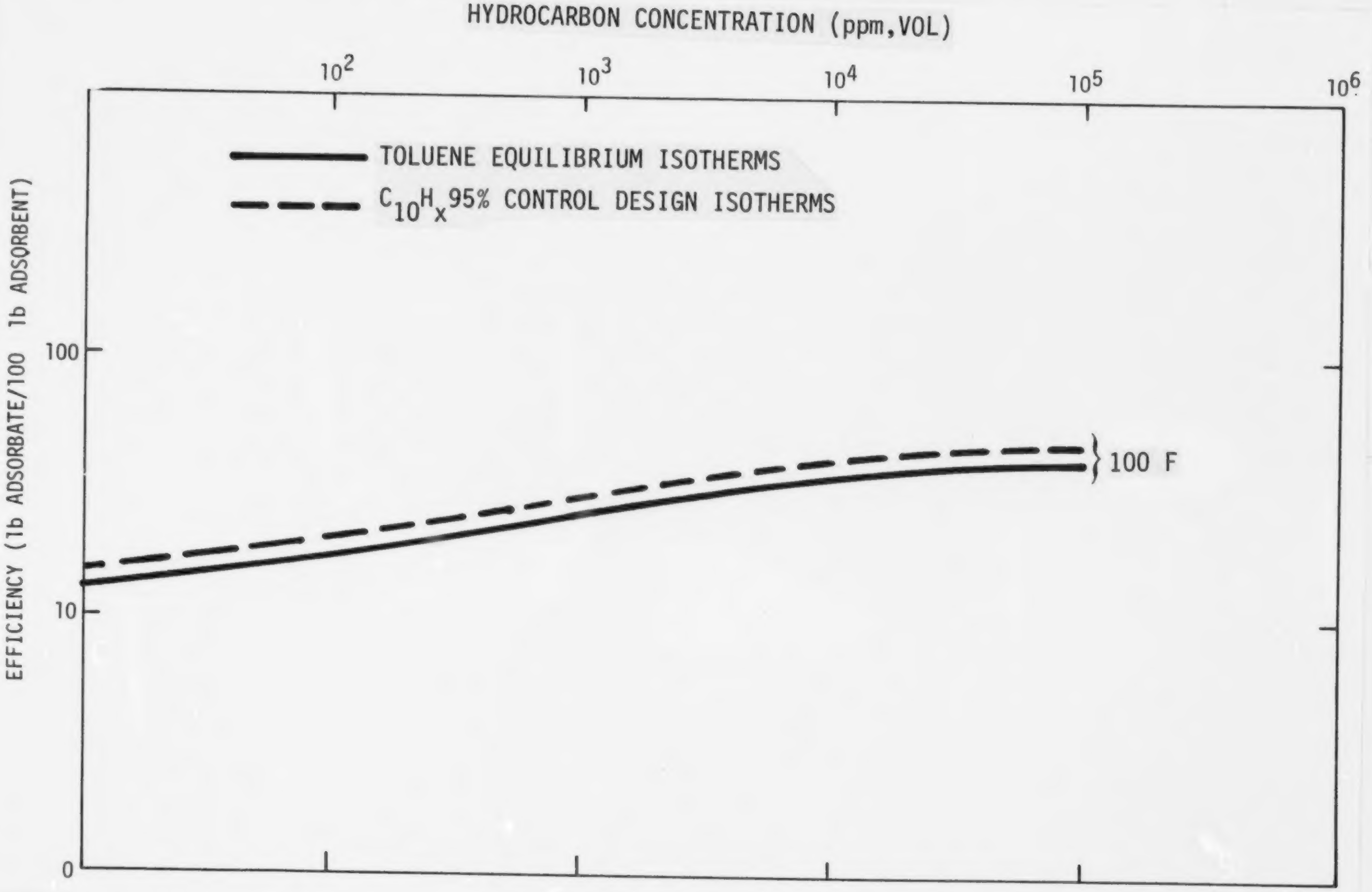

Figure 15. A comparison of $C_{10}H_x$ design and toluene equilibrium adsorption efficiencies for fluidized bed adsorption.

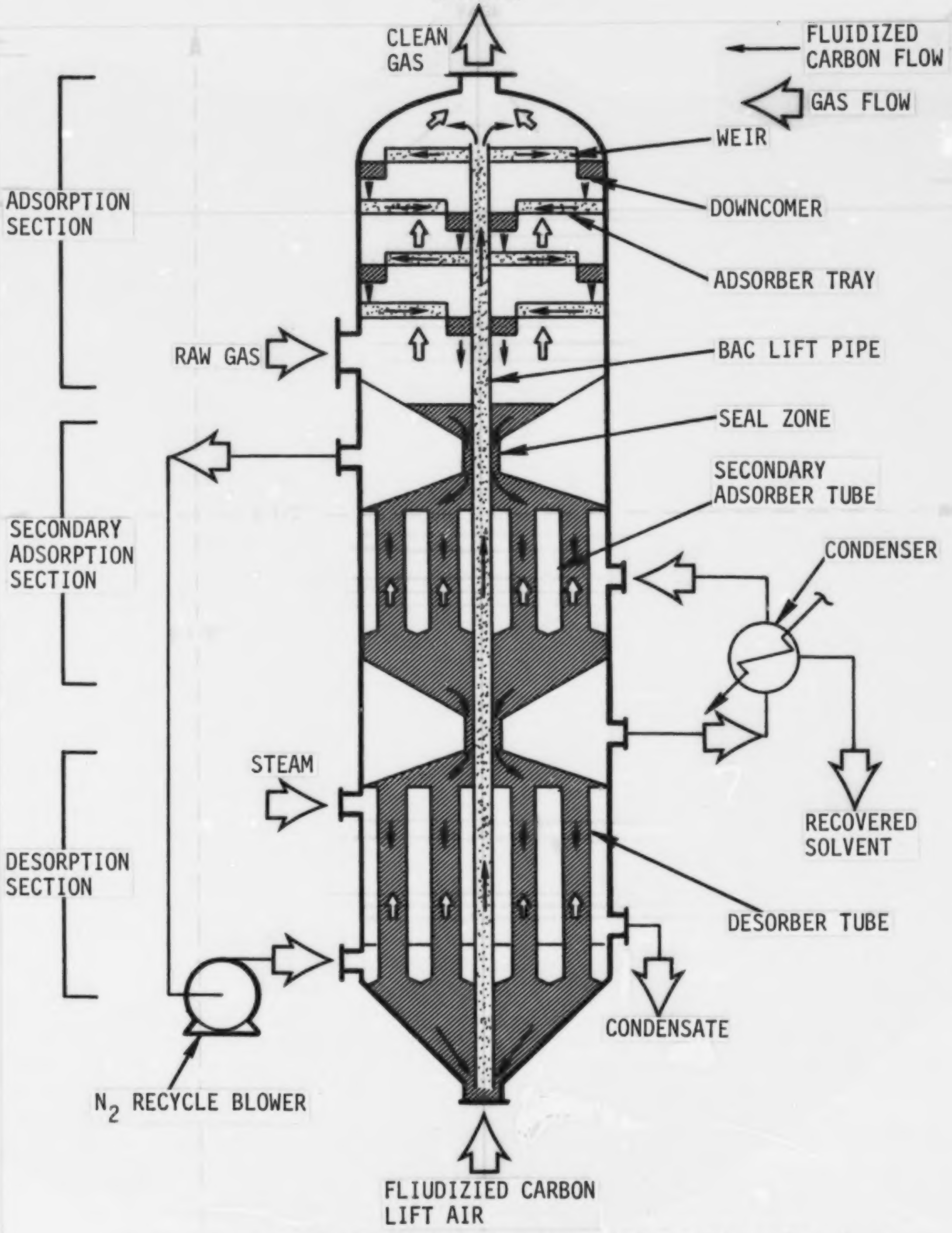
CLEAN GAS
FLUIDIZED CARBON FLOW
GAS FLOW
WEIR
DOWNCOMER
ADSORPTION SECTION
ADSORBER TRAY
RAW GAS
BAC LIFT PIPE
SEAL ZONE
SECONDARY ADSORBER TUBE
SECONDARY ADSORPTION SECTION
CONDENSER
STEAM
RECOVERED SOLVENT
DESORPTION SECTION
DESORBER TUBE
CONDENSATE
N_2 RECYCLE BLOWER
FLIUDIZIED CARBON LIFT AIR

Figure 16. Fluidized bed activated carbon adsorber system configuration.

$$\dot{m}_c = \frac{0.95 \times 5.5 \ \frac{\text{lb solvent}}{\text{min}}}{\frac{35.9 \text{ lb solvent}}{100 \text{ lb carbon}}} = 14.5 \text{ lb/min}$$

From (17) the carbon inventory of the system is 2.5 × 14.5 × 60 = 2180 lb, and the annual replacement rate assumed is 2 percent or 44 lb/year.

Size of Adsorption Tower--

Fluidized bed adsorber size is dictated by the permissible fluidizing velocity. Hence, for 5000 scfm the equipment size for all concentrations is the same. Permissible fluidizing velocity, u_f, is 180 ft/min. From Equation 17 the actual gas flow rate at 100°F is 5385 ft^3/min. The flow cross-sectional area is given by:

$$\frac{\pi D_a^2}{4} = \frac{5385}{180} = 29.9 \text{ ft}^2.$$

Therefore, chamber diameter, $D_a = 6.2$ ft.

Since the carbon saturation profile in the bed remains constant due to the recirculation of freshly regenerated carbon, a low gas/carbon contact time of 0.6 sec is recommended (22).

Total bed depth is (180/60)(0.6)(12) = 21.6 in. $\simeq$ 22 in. Bed depth in any one stage is kept below 6 in. Thus, four stages with a bed depth, X_b, of 5.5 in. per stage is selected.

For X_b bed depth, disengagement heights of X_b between stages and $3\ X_b$ above the top tray are required. For uniform gas distribution through a 6.2 ft diam chamber, a 3.1 ft inlet plenum height, X_p, is recommended.

Hence, the adsorption tower height, $X_A = (4\ X_b + 3\ X_b + 3\ X_b)/12 + X_p = 7.7$ ft.

Given a carbon bulk density ρ_{cb} of 37 lb/ft^3 and a bed expansion factor of 1.2:

- Fluidized carbon bulk density, $\rho_{fc} = 37/1.2 = 31$ lb/ft^3
- Pressure drop through carbon, $\Delta p_c = X_b\ \rho_{fc}/\rho_w$ = 2.7 in. WG
- Distributor pressure drop, $\Delta p_d = 0.3\ \Delta p_c = 0.8$ in. WG.

Hence, for four stages, the adsorption section pressure drop is

$$\Delta p_A = 4(\Delta p_c + \Delta p_d) = 14.0 \text{ in. WG}$$

Desorption Tower Size--

The desorption section is a moving bed. Hence, desorption time is anticipated to be the same as for the fixed bed adsorber. Thus, given $t_{regen} + t_c = 64$ min and $\dot{m}_c = 14.5$ lb carbon/hr, the carbon volume in the desorption section of the tower is:

$$V_c = \frac{\dot{m}_c \times (t_{regen} + t_c)}{\rho_{cb}} = 25 \text{ ft}^3$$

Assuming heat exchanger volume is equal to the carbon volume, the total desorption volume, V_D, is equal to $2 V_c = 50 \text{ ft}^3$. Desorption volume is also equal to $\pi D^2 X_d/4$.

Hence, desorption bed height is given by:

$$X_d = \frac{4 V_D}{\pi D_a^2} = 1.7 \text{ ft}$$

An intermediate sealing section equal to $0.5 X_d$ and a bottom holding hopper equal to X_d are allowed for in calculation.

Thus, desorption tower height, $X_D = X_d + 0.5 X_d + X_d = 4.2$ ft, and system height, $X = X_A + X_D = 4.2 + 7.7 = 11.9$ ft.

Utility Requirements--

As was done in the previous analysis for fixed bed adsorption, values of utility requirements are determined from values reported (17) for a Naphtha system in proportion to gas flow rate for power, and solvent recovered for cooling water and nitrogen. Reference (17) also suggests that desorption steam rate for fluidized beds is 25 percent of that for fixed beds under the same conditions.

The derived requirements are as follows:

- Power: 12.1 kW
- Cooling water: 3.8 tons/hr
- Nitrogen: 48.3 standard ft^3/hr
- Steam: 408 lb/hr.

7.7.3 System Costs (3000 ppm case)

Complexities in the fabrication of a single vessel fluidized bed adsorber make it difficult to estimate equipment cost from vessel sizes. Reference (17) suggests that capital cost of a fluidized bed adsorption system is slightly higher than that for a fixed bed system. Thus, a 5 percent increase in capital cost has been assumed. The ratio of installation cost and maintenance cost to capital cost reported have also been assumed to apply here.

Capital Cost--

1980 fixed bed adsorption system cost	$ 75,700
Installation cost	18,900
Total installed equipment	$ 94,600
Capital cost adjustment	27,900
Total	$122,500

Operating Cost--

Steam cost	$ 9,800
Power cost	4,000
Cooling water cost	800
Nitrogen cost (\$0.006/ft^3 (17))	1,700
Labor cost (0.25 man-years)	4,500
Maintenance cost	3,000
Carbon replacement cost	500
Total annual operating cost	$24,300

Operating Savings--

Annual savings with solvent reuse and fuel value are the same as for the fixed bed adsorber. That is,

- With solvent reuse: $339,100/year
- With solvent as fuel: $109,300/year

7.7.4 System Variations with Concentration

Only operating costs and savings with solvent recovery will change with concentration. Basic equipment costs do not vary significantly with concentration for a constant volume flow. However, differences in initial carbon inventory cost do vary and have been accounted for in capital cost evaluation. The economic evaluation of the fluidized bed adsorption system is summarized under identification numbers X and XI in Table 26.

TABLE 26. COSTS FOR FLUIDIZED BED CARBON ADSORPTION SYSTEMS (BASED ON PRELIMINARY DESIGN)

		Identification number X	Identification number XI
		Resource recovery	
Stack gas emissions concentration (ppm, volume)	Cost item	Recovered solvent reused	Recovered solvent used as fuel
200	Total capital cost ($)	84,800	84,800
	Annual direct cost ($)	+13,000	-3,700
	Payback*	6.5	-
3000	Total capital cost ($)	122,500	122,500
	Annual direct cost ($)	+314,800	+85,000
	Payback*	0.4	1.4
8000	Total capital cost ($)	177,200	177,200
	Annual direct cost ($)	+857,900	+244,600
	Payback*	0.2	0.7
Notes	Negative annual direct cost indicates expenditure; positive annual direct cost indicates earned revenue		

*Payback in years.

7.8 COMPARATIVE ECONOMICS OF APPLICABLE CONTROL TECHNOLOGIES

Payback periods have been used in this study for preliminary comparison of the economics of the applicable control systems. Payback periods for systems operating at the design conditions throughout the year are summarized in Table 27.

Textile mill operating practices necessitate varying the print paste solvent content frequently, causing variations in VOC emissions. To gauge the impact of such practices on the economics of the system, various combinations of operations at 200, 3000 and 8000 ppm emissions for different parts of a year have been considered. The 10 cases chosen, where operation at any one concentration occurs for less than a whole year, are presented in Table 28, as case numbers 4 through 13. Thus, case 6 conditions refer to an operation where the emissions are 3000 ppm for 80 percent, and 8000 ppm for 20 percent of the year. The economic impact is summarized in Table 29, and illustrated in Figures 17 and 18 for systems with solvent reused and used as fuel, respectively. Systems that did not result in paybacks have been excluded.

7.8.1 Interpretation of Table 29 and Figures 17 and 18

The following are interpretations of Table 29 and Figures 17 and 18 for example cases. Consider a fluidized bed adsorption system with recovered solvent used as fuel (identification number XI).

Example 1: Case 4, Table 28--
The textile mill emissions are 200 ppm for 80 percent, and 3000 ppm for 20 percent of the year. The equipment, therefore, must be designed for 3000 ppm.

The total capital cost of the system for 3000 ppm, from Table 27, is $122,500.

Annual direct costs are:

0.8 (-3,700)
+0.2 (+85,000) = $43,000

That is, the annual revenue with the recovered solvent used as fuel is $14,000.

The payback period = 122,500/14,000 = 8.8 years

TABLE 27. SUMMARY OF COSTS AND PAYBACK PERIOD FOR APPLICABLE CONTROL TECHNOLOGIES (BASED ON PRELIMINARY DESIGN)

Stack gas emissions concentration (ppm, volume)	Identification numbers	I	II	III	IV	V	VI	VII	VIII	XI	X	XI
	Control technology	Refrigeration/ condensation		Thermal incineration (stack gas diluted to 3500 ppm, if higher)		Fixed bed catalytic incineration (stack gas diluted to 1580 ppm, if higher)		Fluidized bed catalytic incineration	Fixed bed carbon adsorption (stack gas diluted to 3500 ppm, if higher)		Fluidized bed carbon adsorption	
	Resource recovery / Cost item	Recovered solvent reused	Recovered solvent used as fuel	No heat recovery	Heat recovery	No heat recovery	Heat recovery	Heat recovery	Recovered solvent reused	Recovered solvent used as fuel	Recovered solvent reused	Recovered solvent used as fuel
200	Total capital cost ($)	505,100	505,100	73,200	165,600	105,800	185,700	214,300	140,000	140,000	94,800	84,800
	Annual direct cost ($)	-75,200	-91,500	-103,000	-21,100	-76,600	-34,300	-37,800	-7,200	-23,900	+13,000	-3,700
	Payback	-	-	-	-	-	-	-	-	-	6.5	-
3000	Total capital cost ($)	505,100	505,100	71,300	174,800	144,600	240,400	224,800	140,000	140,000	122,500	122,500
	Annual direct cost ($)	+250,800	+10,800	-15,500	+80,600	-32,000	+60,200	+63.300	+262,500	+32,700	+314,800	+85,000
	Payback	2.0	4.7	-	2.2	-	4.0	3.6	0.5	4.3	0.4	1.4
8000	Total capital cost ($)	505,100	505,100	106,000	324,100	249,400	469,300	256,500	259,600	259,600	177,200	177,200
	Annual direct cost ($)	+820,800	+195,800	-19,500	+249,100	-46,600	+211,500	+242,500	+734,300	+121,000	+857,900	+244,600
	Payback	0.6	2.6	-	1.3	-	2.2	1.1	0.4	2.1	0.2	0.7

Notes: Negative annual direct cost indicates expenditure; positive annual direct cost indicates earned revenues.

TABLE 27. SUMMARY OF COSTS AND PAYBACK PERIOD FOR APPLICABLE CONTROL TECHNOLOGIES (BASED ON PRELIMINARY DESIGN)

Stack gas emissions concentration (ppm, volume)	Identification numbers	I	II	III	IV	V	VI	VII	VIII	XI	X	XI
	Control technology	Refrigeration/ condensation		Thermal incineration (stack gas diluted to 3500 ppm, if higher)		Fixed bed catalytic incineration (stack gas diluted to 1580 ppm, if higher)		Fluidized bed catalytic incineration	Fixed bed carbon adsorption (stack gas diluted to 3500 ppm, if higher)		Fluidized bed carbon adsorption	
	Resource recovery Cost item	Recovered solvent reused	Recovered solvent used as fuel	No heat recovery	Heat recovery	No heat recovery	Heat recovery	Heat recovery	Recovered solvent reused	Recovered solvent used as fuel	Recovered solvent reused	Recovered solvent used as fuel
200	Total capital cost ($)	505,100	505,100	73,200	165,600	105,800	185,700	214,300	140,000	140,000	84,800	84,800
	Annual direct cost ($)	-75,200	-91,500	-103,000	-21,100	-76,600	-34,300	-37,800	-7,200	-23,900	+13,000	-3,700
	Payback	-	-	-	-	-	-	-	-	-	6.5	-
3000	Total capital cost ($)	505,100	505,100	71,300	174,800	144,600	240,400	224,800	140,000	140,000	122,500	122,500
	Annual direct cost ($)	+250,800	+10,800	-15,500	+80,600	-32,000	+60,200	+63.300	+262,500	+32,700	+314,800	+85,000
	Payback	2.0	4.7	-	2.2	-	4.0	3.8	0.5	4.3	0.4	1.4
8000	Total capital cost ($)	505,100	505,100	106,000	324,100	249,400	469,300	256,500	259,600	259,600	177,200	177,200
	Annual direct cost ($)	+820,800	+195,800	-19,500	+249,100	-36,600	+211,500	+242,500	+734,300	+121,000	+857,900	+244,600
	Payback	0.6	2.6	-	1.3	-	2.2	1.1	0.4	2.1	0.2	0.7

Notes: Negative annual direct cost indicates expenditure; positive annual direct cost indicates earned revenues.

TABLE 28. CHANGES IN DURATION OF EMISSIONS AT DIFFERENT CONCENTRATIONS

Description	Case number for Figures 17 and 18	Emissions concentration for percent of year		
		200 ppm (%)	3000 ppm (%)	8000 ppm (%)
Operation at single concentration level	1	100	0	0
	2	0	100	0
	3	0	0	100
Predominant concentration level with overlap into one other level	4	80	20	0
	5	20	80	0
	6	0	80	20
	7	0	20	80
Operation at two concentration levels, one slightly dominant	8	60	40	0
	9	60	0	40
	10	40	0	60
	11	0	60	40
	12	0	40	60
Operation at all levels equally	13	33	34	33

TABLE 29. EFFECT OF CHANGES IN DURATION OF EMISSIONS AT DIFFERENT CONCENTRATIONS ON THE ECONOMICS OF SELECTED CONTROL TECHNOLOGIES

Case number	Cost item	Identification numbers I	II	IV	VI	VII	VIII	IX	X	XI	Comments
1	Capital	505,100	505,100	165,600	185,700	214,300	140,000	140,000	84,800	84,800	100% low concentration
	Annual direct	-75,200	-91,500	-21,100	-34,300	-37,800	-7,200	-23,900	+13,000	-3,700	
	Payback	-	-	-	-	-	-	-	6.5	-	
2	Capital	505,100	505,100	174,800	240,400	224,800	140,000	140,000	122,500	122,500	100% medium concentration
	Annual direct	+250,800	+10,800	+80,600	+60,200	+63,300	+262,500	+32,700	+314,800	+85,000	
	Payback	2.0	47	2.2	4.0	3.6	0.5	4.3	0.4	1.4	
3	Capital	505,100	505,100	324,100	469,300	256,500	259,600	259,600	177,200	177,200	100% high concentration
	Annual direct	+820,800	+195,800	+249,100	+211,500	+242,500	+734,300	+121,000	+857,900	+244,600	
	Payback	0.6	2.6	1.3	2.2	1.1	0.4	2.1	0.2	0.7	
4	Capital	505,100	505,100	174,800	240,400	224,800	140,000	140,000	122,500	122,500	Predominantly low some medium concentrations
	Annual direct	-10,000	-71,000	-800	-15,400	-17,600	+46,700	-12,000	+73,400	+14,000	
	Payback	-	-	-	-	-	3.0	-	1.7	8.8	
5	Capital	505,100	505,100	174,800	240,400	224,800	140,000	140,000	122,500	122,500	Predominantly medium some low concentrations
	Annual direct	185,600	-9,700	+60,300	+41,300	+43,100	+208,600	+21,400	+254,400	+67,300	
	Payback	2.7	-	2.9	5.8	5.2	0.7	6.5	0.5	1.8	
6	Capital	505,100	505,100	324,100	469,300	256,500	259,600	259,600	177,200	177,200	Predominantly medium some high concentrations
	Annual direct	+364,800	+47,800	+114,300	+90,500	+99,100	+356,900	+50,400	+423,400	+116,900	
	Payback	1.4	10.6	2.8	5.2	2.6	0.7	5.2	0.4	1.5	
7	Capital	505,100	505,100	324,100	469,300	256,500	259,600	259,600	177,200	177,200	Predominantly high some medium concentrations
	Annual direct	+706,800	+158,800	+215,400	+181,200	+206,700	+639,900	+103,300	+749,300	212,700	
	Payback	0.7	3.2	1.5	2.6	1.2	0.4	2.5	0.2	0.8	

(Continued)

TABLE 29 (continued).

Case number	Cost item	I	II	IV	VI	VII	VIII	IX	X	XI	Comments
		Identification numbers									
8	Capital	505,100	505,100	174,800	240,400	224,800	140,000	140,000	122,500	122,500	All operations at low and medium concentrations
	Annual direct	+55,200	-50,600	+19,600	+3,500	+2,600	+100,700	-1,300	+126,700	+31,800	
	Payback	9.2	-	8.9	69	86	1.4	-	1.0	3.9	
9	Capital	505,100	505,100	324,100	469,300	256,500	259,600	259,600	177,200	177,200	All operations at low and high concentrations
	Annual direct	+283,200	+23,400	+87,000	+64,000	+74,300	+289,400	+34,100	+351,000	+95,600	
	Payback	1.8	22	3.7	7.3	3.5	0.9	7.6	0.5	1.9	
10	Capital	505,100	505,100	324,100	469,300	256,500	259,600	259,600	177,200	177,200	All operations at low and high concentrations
	Annual direct	+462,400	+80,900	+141,000	+113,200	+130,400	+437,700	+63,000	+519,900	+145,300	
	Payback	1.1	6.2	2.3	4.1	2.0	0.6	4.1	0.3	1.2	
11	Capital	505,100	505,100	324,100	469,300	256,500	259,600	259,600	177,200	177,200	All operations at medium and high concentrations
	Annual direct	+478,800	+84,800	+148,000	+120,700	+135,000	+451,200	+68,000	+532,000	+148,800	
	Payback	1.1	6.0	2.2	3.9	1.9	0.6	3.8	0.3	1.2	
12	Capital	505,100	505,100	324,100	469,300	256,500	259,600	259,600	177,200	177,200	All operations at medium and high concentrations
	Annual direct	+592,800	+121,800	+181,700	+151,000	+170,800	+545,600	+85,700	+640,700	+180,800	
	Payback	0.9	4.1	1.8	3.1	1.5	0.5	3.0	0.3	1.0	
13	Capital	505,100	505,100	324,100	469,300	256,500	259,600	259,600	177,200	177,200	Operations at low, medium and high concentrations
	Annual direct	+331,300	+38,100	+102,600	+78,900	+89,100	+329,200	+43,200	+394,400	+108,400	
	Payback	1.5	13	3.2	5.9	2.9	0.8	6.0	0.4	1.6	

Note: Positive annual direct costs indicate revenue. Negative annual direct costs indicate expenditures.
See Table 27 for identification numbers.
See Table 28 for case numbers.

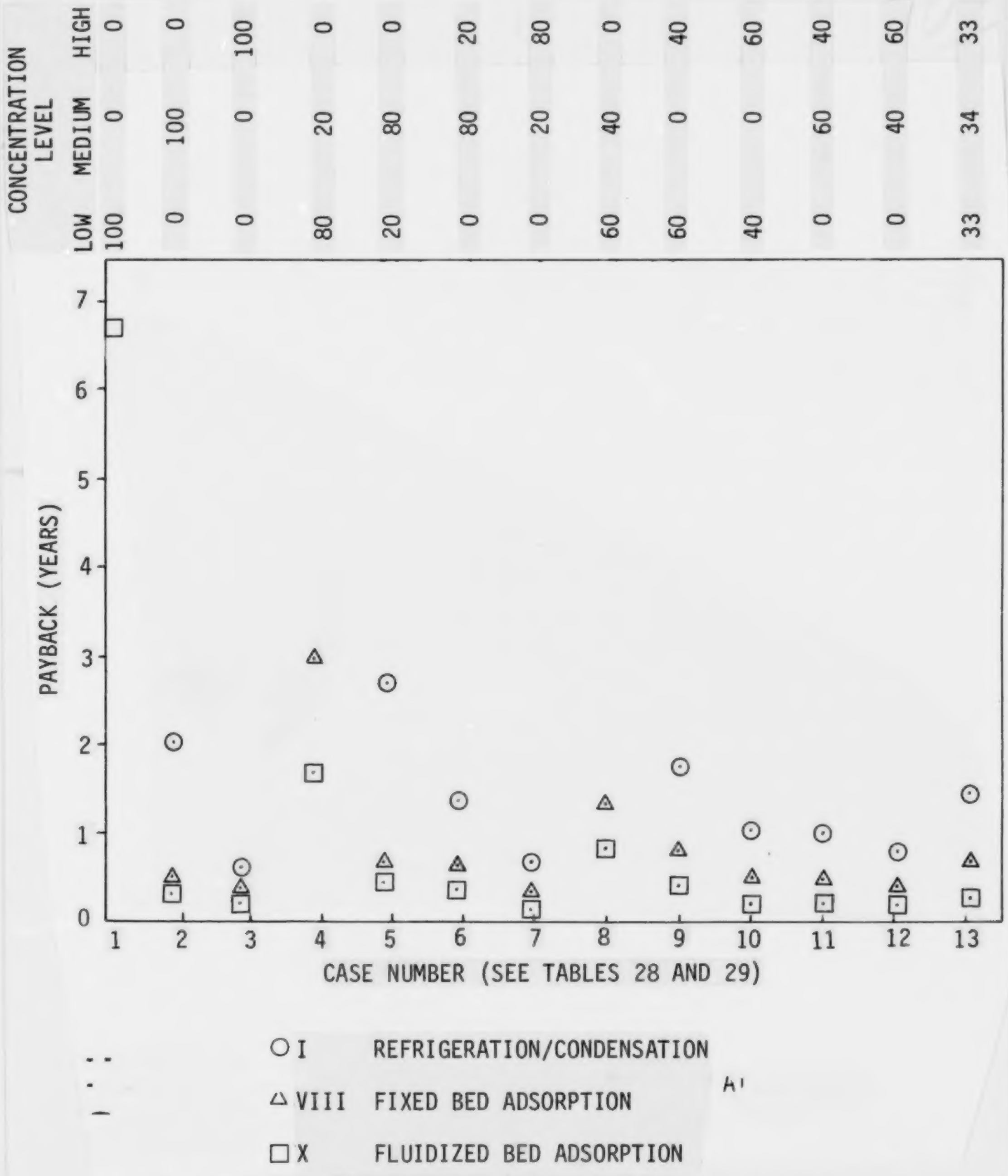

Figure 17. Variation of payback period with changes in duration of emissions at different concentrations: recovered solvent reused.

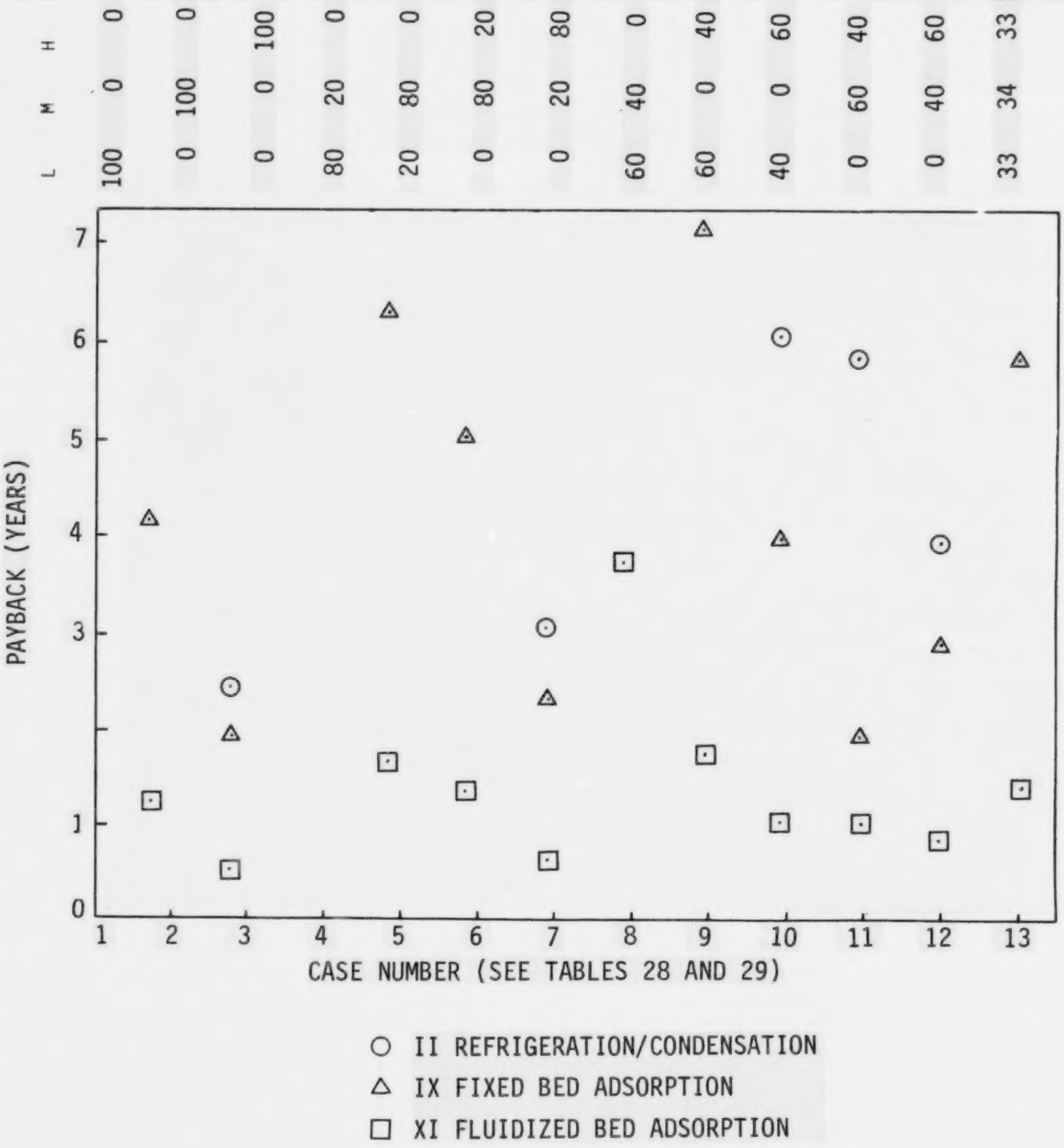

Figure 18. Variation of payback period with changes in duration of emissions at different concentrations: recovered solvent used as fuel.

Example 2: Case 8, Table 28--
Emissions durations are:

- 60 percent of the year at 200 ppm
- 40 percent of the year at 3000 ppm.

The total capital cost of a fluidized bed adsorption system designed for 3000 ppm, from Table 27, is $122,500.

Annual direct costs with recovered solvent as fuel are:

$$\begin{aligned} &0.6\ (-3{,}700) \\ +&0.4\ (+85{,}000) = \quad \underline{\$32{,}800} \end{aligned}$$

Payback period = 122,500/32,800 = 3.7 years.

Example 3: Case 12, Table 28--
Emissions durations are:

- 40 percent of the year at 3000 ppm
- 60 percent of the year at 8000 ppm.

The total capital cost of a fluidized bed adsorption system designed for 8000 ppm, from Table 27, is $177,200.

Annual direct costs with recovered solvent as fuel are:

$$\begin{aligned} &0.4\ (+85{,}000) \\ +&0.6\ (+244{,}600) = \quad \underline{\$180{,}800} \end{aligned}$$

Payback period = 177,200/180,800 = 1.0 years

Thus, the length of periods over which the emissions are low, medium or high impacts the economic viability of a control device. Figure 17 is a graphical depiction of the variations in payback period when any solvent recovered has reuse value and Figure 18 is the same when the solvent has only a fuel value. The purpose of this study is to identify economically viable control options. Given the varying nature of the textile industry, those controls that show the best payback throughout the range of concentration changes can be interpreted as having the best potential for economic viability. When payback periods are within 25 percent of each other, alternate systems are considered competitive.

7.8.2 Preliminary Conclusions

The results of the preliminary analysis of emissions control technologies, reported in terms of payback in Table 29 are summarized below:

- The only control technology with projected paybacks for all concentration conditions considered is fluidized bed adsorption with recovered solvent reused. No other technology projects a payback for 100 percent operation at the low concentration level (200 ppm).

- As long as operation at low level exhaust concentration is not more than 80 percent, fixed bed adsorption and refrigeration/condensation both with reuse of recovered solvent also project reasonable payback periods for all other concentration combinations.

- Considering the top three technologies ranked by average payback, the payback period for fixed bed adsorption and refrigeration/condensation is approximately twice and three and one-half times, respectively, that of fluidized bed adsorption (all with recovered solvent reuse).

- The fourth ranked technology overall is fluidized bed adsorption with recovered solvent used as fuel. This technology actually projects shorter paybacks than refrigeration/condensation with solvent reuse if operation never occurs at high exhaust concentration (8000 ppm).

- Fluidized bed catalytic incineration and thermal incineration both with heat recovery are closely ranked technologies. Overall the former has the fifth best average payback, although the latter projects better paybacks when high concentrations never occur during operation.

- The final three technologies considered, in order of average payback are fixed bed catalytic incineration with heat recovery, fixed bed adsorption with recovered solvent used as fuel, and refrigeration/condensation with recovered solvent used as fuel. In order for these techniques to realize a reasonable payback, concentrations must be in the medium (3000 ppm) to high (8000 ppm) concentration range better than 50 percent of the time. In order for refrigeration/condensation with solvent as fuel to project a payback of less than 10 years, better than 20 percent operation in the high range is required if the remaining operation is at medium concentration, or better than 50 percent is required in the high range if the remainder is at low concentration.

- Because the value of recovered solvent is three times greater with reuse than as fuel, systems with solvent recovery have better paybacks with reuse.
- Fluidized bed adsorption has a shorter payback period than fixed bed adsorption despite nearly equal capital costs, due to the lower utility and maintenance costs of the former.

SECTION 8

DETAILED DESIGN AND COSTING OF SELECTED CONTROL TECHNOLOGIES

The design and costs for the adsorber and catalytic incineration systems were developed in sufficient detail in the previous section to allow a reasonable cost-benefit evaluation of these technologies. The refrigeration/condensation aerosol removal system and the thermal incineration system designs are considered in more detail in this section in order to provide a more comprehensive basis for final ranking of applicable emission control technologies. Analyses for the 200 ppm case are presented since the overall economic viability of a system is greatly influenced by the system performance at low emission concentration levels. System analyses for the 3000 and 8000 ppm cases have been performed to the same degree of detail and the results are reported in summary in Table 31.

The operating conditions used here are the same as used in the preliminary evaluation. That is, for 5000 scfm of 200 ppm VOC concentration stack gas at T_1 = 300°F and 6 percent relative humidity, the inlet gas conditions are

$\dot{m} = 343.1$ lb/min

$\dot{m}_a = 278.6$ lb/min

$\dot{m}_w = 64.1$ lb/min

$\dot{m}_s = 0.37$ lb/min

$c_p = 0.281$ Btu/lb°F

8.1 DETAILED DESIGN OF REFRIGERATION/CONDENSATION AEROSOL REMOVAL SYSTEM

8.1.1 System Configuration

The refrigeration/condensation system configuration for low concentration emissions is presented in Figure 19 and is basically the same as that for medium concentration emissions which was used in the preliminary system evaluation.

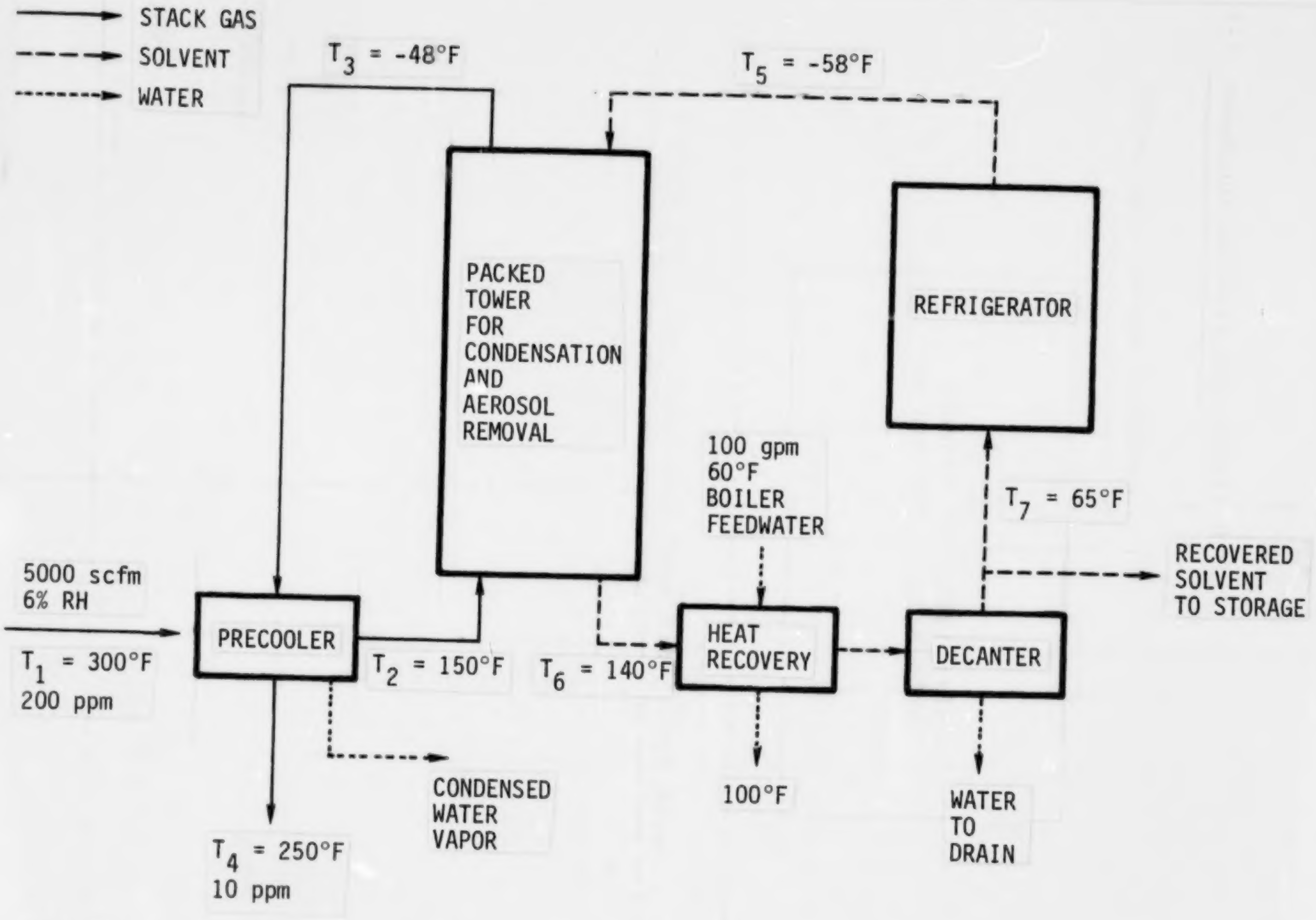
STACK GAS
SOLVENT
WATER
T_3 = -48°F
T_5 = -58°F
PACKED TOWER FOR CONDENSATION AND AEROSOL REMOVAL
REFRIGERATOR
100 gpm
60°F
BOILER FEEDWATER
T_7 = 65°F
RECOVERED SOLVENT TO STORAGE
5000 scfm
6% RH
T_1 = 300°F
200 ppm
PRECOOLER
T_2 = 150°F
T_6 = 140°F
HEAT RECOVERY
DECANTER
CONDENSED WATER VAPOR
100°F
WATER TO DRAIN
T_4 = 250°F
10 ppm

Figure 19. System configuration for refrigeration/condensation aerosol removal control technology, detailed design (200 ppm case).

8.1.2 Detailed System Design (200 ppm Case)

Precooler --

From Equations 3 and 4 and Figure 7 the condenser outlet temperature, T_3, required to achieve 95 percent control with 98 percent efficient aerosol removal is -48°F. The condenser outlet conditions are therefore:

$\dot{m} = 279.8$ lb/min

$\dot{m}_a = 278.6$ lb/min

$\dot{m}_w = 1.2$ lb/min

$\dot{m}_s = 0.01$ lb/min

$c_p = 0.239$ Btu/lb°F

$\rho = 0.0968$ lb/ft^3

Given a minimum temperature difference of 50°F for gas-to-gas heat exchangers, the temperature, T_4, and c_p for the purified gas leaving the precooler are 250°F and 0.242 Btu/lb°F, respectively. These conditions minimize the size of the refigeration system required to achieve the necessary cooling.

As was done for the preliminary system evaluation, the cooling capacity available from the condenser outlet gas is first determined from Equation 5. Using the average value of c_p,

$$H_{\text{precooler}} = 279.8 \times 0.240 \times (250 - (-48))$$
$$= 20{,}010 \text{ Btu/min}$$

This value is then used in Equation 7 to determine, by iteration, the rate at which moisture from the stack gas condenses in the precooler and the precooled stack gas (condenser inlet) temperature, T_2. Since no special provisions for aerosol removal are included in the precooler design, 50 percent of the condensed moisture is assumed to be carried on to the condenser as aerosol. A 10 percent loss of cooling capacity due to thermal exchange with the environment is also assumed. The resulting conditions at the condenser inlet are:

$T_2 = 150$°F

$\dot{m} = 341.6$ lb/min

$c_p = 0.280$ Btu/lb°F

$\rho = 0.0596$ lb/ft^3

The rate of condensed water removal from the precooler is approximately 1.5 lb/min.

For the precooler capacity of 0.90 × 20,010 = 18,010 Btu/min, the required surface area of the exchanger is

$$A_{precooler} = (18{,}010 \text{ Btu/min} \times 60 \text{ min/hr})/$$
$$(3.5 \text{ Btu/hr ft}^2\text{°F} \times 108\text{°F})$$
$$= 2860 \text{ ft}^2$$

Condensation/Aerosol Removal--

In the condenser, the stack gas must be cooled to -48°F. Since the water vapor content of the stack gas exceeded the saturation level in the precooler, water will continue to condense out of the gas stream as it passes through the condenser. With 98 percent efficient aerosol removal, 61.3 lb water/min will be removed from the stack gas. The solvent vapor will only reach its saturation level at 16°F. As the stack gas is cooled from 16°F to -48°F, approximately 0.36 lb solvent/min will be condensed and removed from the stack gas.

The cooling load of the condenser includes the energy required to cool the stack gas air and uncondensed vapors from 150°F to -48°F, that required to cool water and solvent vapors from 150°F to the temperature at which they condense (assume the mean temperature between saturation and outlet temperatures) and the heat of condensation of the condensed water and solvent. Thus,

$$H_{condenser} = 278.6 \times 0.240 \times (150 - (-48))$$
$$+ 61.1 \times 0.446 \times (150 - 51)$$
$$+ 0.36 \times 0.37 \times (150 - (-16))$$
$$+ 61.1 \times 1065 + 0.36 \times 120$$
$$= 81{,}100 \text{ Btu/min}$$

Given a refrigerated solvent inlet temperature, T_5, of -58°F and a maximum temperature for the liquid leaving the condensation/aerosol removal unit, T_6, of 140°F, the minimum refrigerated solvent flow rate is determined from a thermal balance. The heat removed from the gas stream will raise the temperature of the recovered condensed water and solvent from their average condensing temperatures as well as the temperature of the refrigerated solvent. Assuming a 10 percent thermal loss from the unit

$$H_{condenser} = 0.90 \sum_{components} \dot{m} \, c_p \, \Delta T \qquad (28)$$

$$= 0.90 \times \Big[\dot{m}_{s,r} \times 0.5 \times (140 - (-58)) + 61.3 \times 1.00 \times (140 - 51) + 0.36 \times 0.5 \times (140 - (-16))\Big]$$

so that the refrigerated solvent recirculation rate is

$$\dot{m}_{s,r} = 81{,}100/0.90 - 5456 - 28)/99.0$$
$$= 855 \text{ lb/min}$$

A packed tower with refrigerated solvent liquid phase is appropriate for this application since it can serve the dual functions of condensing the air stream vapors and removing the resulting aerosols. The refrigerated liquid solvent is introduced at the top of the tower and presents a large surface of liquid film to the countercurrent gas flow as it runs down and wets the packing material. Gas enters at the bottom of the tower and is cooled by direct contact with the liquid as it flows upward through the packing. As aerosols are formed, they are transferred to the liquid solvent stream.

The tower diameter required to provide the desired degree of emissions control is determined from parametric relationships presented in reference (8). To insure sufficient wetting of the packing without flooding the tower, the tower cross sectional area, A_{XT}, is selected so that the gas velocity is 55 percent of the flooding velocity and the refrigerated solvent flow rate, $\dot{m}_{s,r}$, is greater than 13.4 A_{XT}. With 1 in. Raschig rings as packing the packed tower diameter for the given flow conditions is

$$D_{PT} = 7.3 \text{ ft}$$

Tower volume is given by (23)

$$V_{PT} = \frac{H_{Condenser}}{U_V \times LMTD}$$

where U_V, the volumetric heat transfer coefficient for direct contact heat exchange is 1340 Btu/hr-ft^3-°F (18). Thus

$$V_{PT} = \frac{81{,}100 \times 60}{1340 \times 10} = 363 \text{ ft}^3$$

and the tower height

$$X_{PT} = \frac{V_{PT}}{\frac{\pi D_{PT}^2}{4}} = \frac{363 \times 4}{\pi \times (7.3)^2} = 8.7 \text{ ft}$$

Waste Heat Recovery--

For maximum energy recovery the condensed water and VOC as well as the recirculating liquid VOC are cooled from 140°F to 65°F by 100 gpm of cooling water entering at 60°F and leaving at 100°F. With a 10 percent heat loss from the exchanger the waste heat recovered is determined according to Equation 11 to be

$$H_{WH} = 0.90 \times \left[(855 + 0.36) \times 0.5 + 61.3 \times 1.00\right] \times (140 - 65) = 33{,}000 \text{ Btu/min}$$

and the required exchanger surface area is

$$A_{\text{waste heat recovery}} = (33{,}000 \text{ Btu/min} \times 60 \text{ min/hr})/(12 \text{ Btu/hr ft}^2\text{°F} \times 17\text{°F}) = 9700 \text{ ft}^2$$

Refrigeration Heat Load--

The cooled liquids are decanted, the water is drained, the recovered VOC is sent to storage and 855 lb/min of VOC is refrigerated from 65 to -58°F before reentering the condenser. Equation 10 is used to determine the amount of refrigeration required by the system. Thus

$$H_{ref} = 855 \times 0.5 \times (65 - (-58)) = 52{,}600 \text{ Btu/min}$$

which converts to

$$H'_{ref} = 52{,}600/200 = 263 \text{ tons}$$

Power Requirement--

With 1.5 kW/ton power consumption for refrigeration the annual cost of power for the illustrated system will be

$$263 \text{ tons} \times 1.5 \frac{\text{kW}}{\text{ton}} \times 6000 \frac{\text{hr}}{\text{year}} \times \frac{\$0.055}{\text{kWh}} = \$130{,}200/\text{year}$$

Operating Savings--

The annual value of the waste heat recovered is

$$33{,}000 \frac{\text{Btu}}{\text{min}} \times 60 \frac{\text{min}}{\text{hr}} \times 6000 \frac{\text{hr}}{\text{year}} \times \frac{\$2.9}{10^6 \text{ Btu}} = \$34{,}500/\text{year}$$

If 90 percent of the solvent is recovered and reused its annual value is

$$0.90 \times 0.36\ \frac{\text{lb}}{\text{min}} \times 60\ \frac{\text{min}}{\text{hr}} \times 6000\ \frac{\text{hr}}{\text{year}} \times \frac{\$0.18}{\text{lb}} = \$21{,}000/\text{year}$$

If 90 percent of the solvent is recovered and burned as fuel, its annual value is

$$0.90 \times 0.36\ \frac{\text{lb}}{\text{min}} \times 60\ \frac{\text{min}}{\text{hr}} \times 6000\ \frac{\text{hr}}{\text{year}} \times 19{,}930\ \frac{\text{Btu}}{\text{lb}}$$

$$\times\ \frac{\$2.9}{10^6\ \text{Btu}} = \$6{,}700/\text{year}$$

8.1.3 Optimization of Refrigeration/Condensation Aerosol Removal System

System optimization implies maximizing the return on money invested in the emissions control system. The size and, therefore, cost of the packed tower for condensation and aerosol removal varies very little with stack gas concentration. Also the total amount of cooling required to reduce the stack gas temperature from 300°F to the appropriate condenser exhaust temperature varies little, from 476 to 454 tons as the concentration increases from 200 to 8000 ppm.

Figure 20 presents the annual costs of a system in which all of the cooling is provided by refrigerated solvent recirculating through the packed tower; i.e., no precooler or waste heat recovery is employed. Since the refrigeration capacity does not vary significantly with stack gas concentration, the capital cost of $535,000 is for a system designed for the minimum concentration level of 200 ppm (and, therefore, maximum cooling requirement). The plots of annual direct costs versus concentration demonstrate that operating savings resulting from use of recovered solvent as fuel are not sufficient in themselves to balance the annual operating costs of the system at any concentration level. On the other hand, if the recovered solvent can be reused, operating costs and savings cancel out at approximately 2800 ppm. If an annual return of 15 percent over 10 years is required for a system to be economically viable, then the annual return for this system must be $106,600/year. Thus the mean stack gas concentration should be at least 3700 ppm with recovered solvent reused.

An alternate system would include a precooler exchanger as described in the previous analysis. A maximum capacity precooler would reduce the refrigeration system load and, therefore, the annual refrigeration power cost by approximately 10 percent. The capital cost of the system increases when the precooler is included, as does the annual cost of maintenance. The operating savings remain the same. The costs versus concentration of a

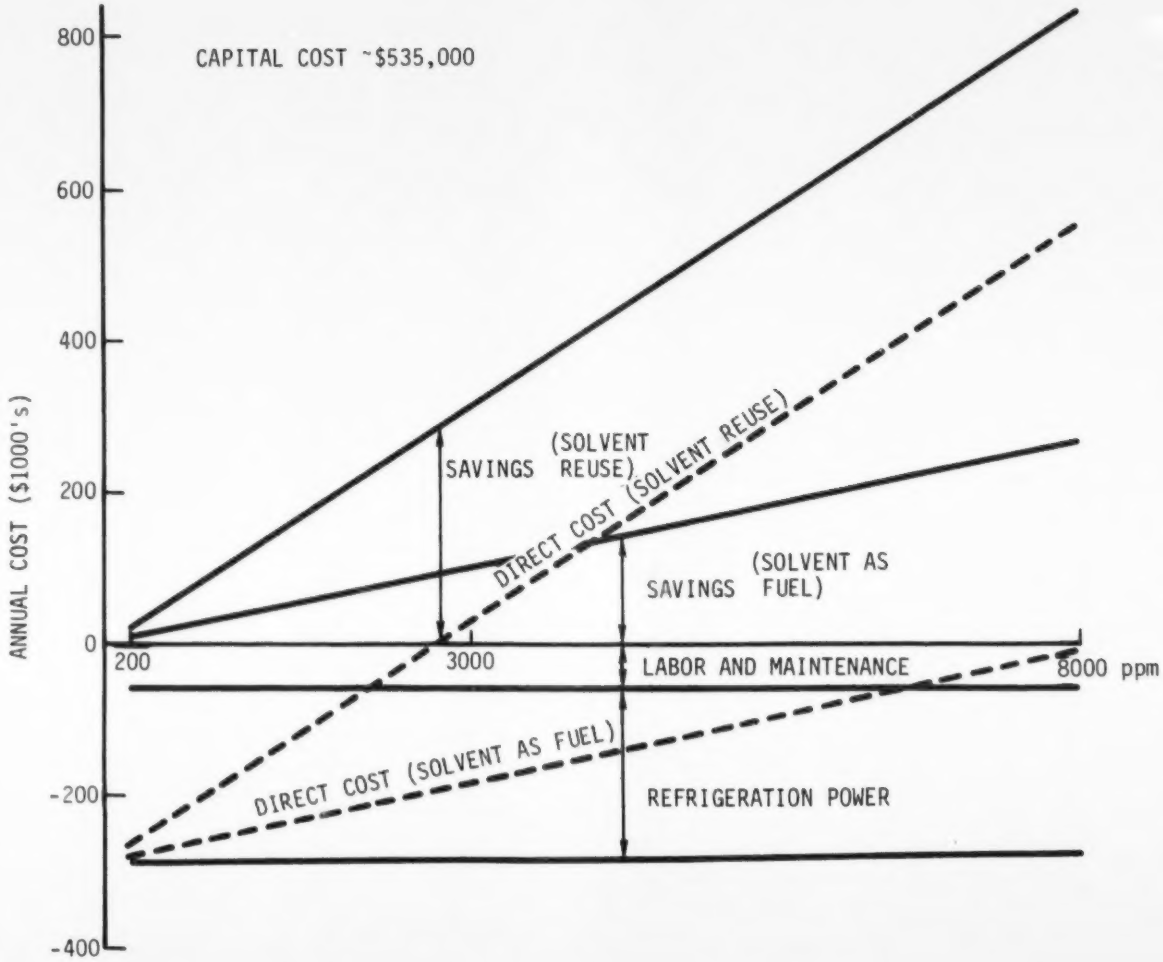

Figure 20. Annual costs versus concentration for refrigeration/ condensation system without precooler or waste heat recovery.

refrigeration/condensation system with a maximum capacity (T_2 = 250°F) precooler are presented in Figure 21. Given the same annual return requirements as for the above system configuration, the zero direct annual cost concentration is again approximately 3700 ppm. Thus, the reduction in operating costs obtained by use of the precooler is counterbalanced by the increase in required annual return resulting from the increase in system capital cost. A better balance could conceivably be derived by optimizing the size of the precooler and, therefore, capital cost with respect to operating cost.

Additional improvement in annual direct cost versus capital investment could be obtained by installing a waste heat recovery exchanger between the packed tower solvent discharge and the refrigeration system. In order for operating savings to be generated by this device, an appropriate use for the recovered thermal energy must be identified. Thermal energy at a maximum temperature of 290°F is available from a system without a precooler and at approximately 140°F from a system with a precooler. Two potential applications for the waste heat recovered here are dryer air preheating and boiler feedwater preheating.

As an example of waste heat exchanger selection for optimum return on investment assume a textile finishing plant has 100 gpm boiler feedwater at 60°F to collect the waste heat from packed tower solvent discharging at 140°F. The waste heat exchanger size will depend on the rate of solvent flow and the desired boiler feedwater outlet temperature. If the full range of concentrations is expected during operation, the exchanger should be sized for the least solvent flow rate (which occurs at 200 ppm). The exchanger will then have a greater surface area than is required for other concentrations. The impact of exchanger design is presented in Figure 22 in curves of capital cost and annual direct cost versus waste heat recovery exchanger size. A heat transfer coefficient of 13 Btu/hr-ft^2-°F has been assumed.

Although the refrigeration system capacity and therefore cost decreases with increasing waste heat exchanger capacity, the exchanger cost increases more rapidly so that the system capital cost increases with exchange size. Annual labor costs and operating savings due to solvent reuse remain the same for a given operating concentration but the reduction in power costs and increase in the value waste heat recovered are much greater than the increased maintenance cost so that annual return increases with exchanger capacity. An annual return is not actually generated unless the mean operating concentration is over 1400 ppm with a maximum capacity exchanger. For smaller exchangers even greater mean concentrations are required to generate revenue.

Given a mean stack gas concentration of 3000 ppm the optimum recovery exchange size is approximately 7000 ft^2 where the ratio of direct cost to capital cost is greatest at 0.30.

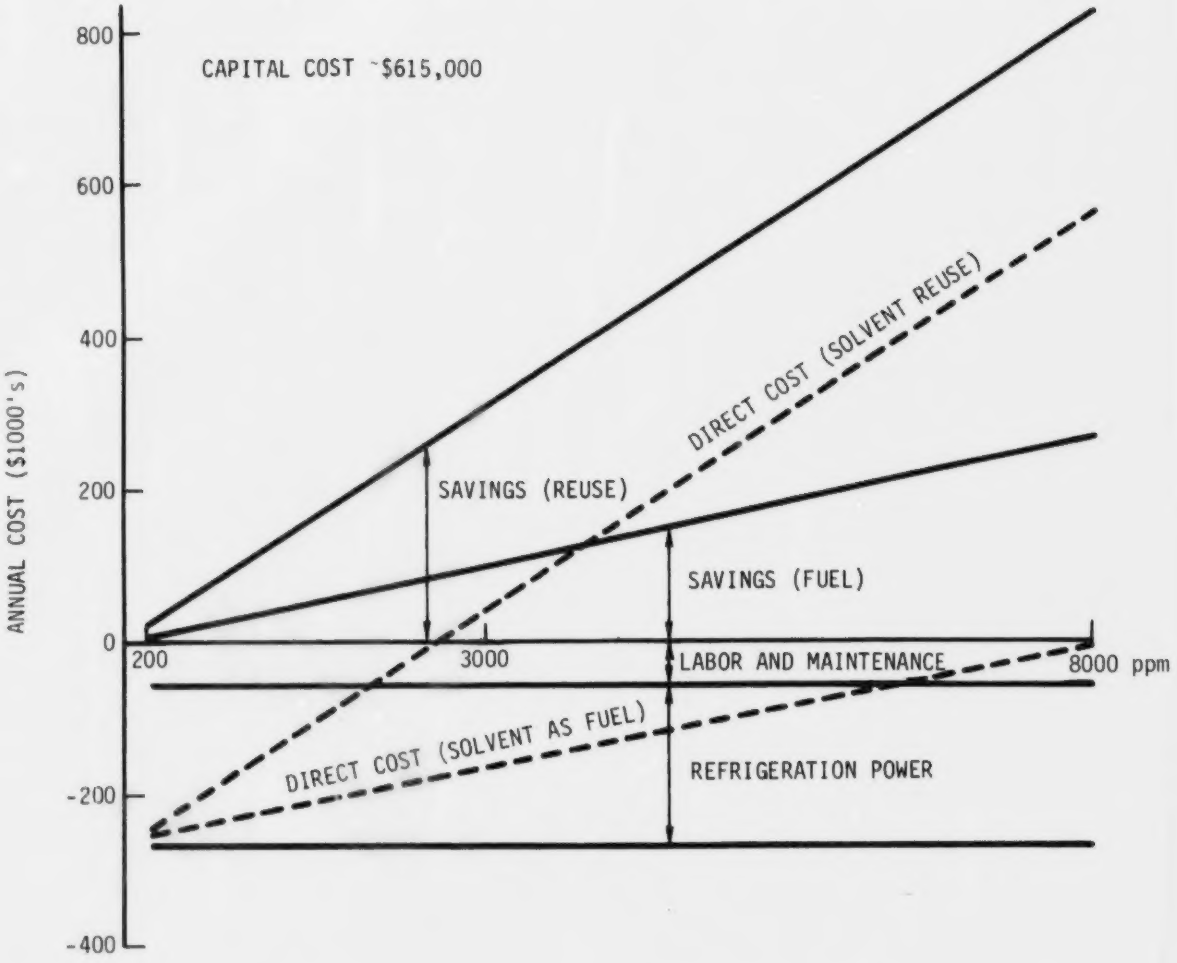

Figure 21. Annual costs versus concentration for refrigeration/ condensation system with precooler.

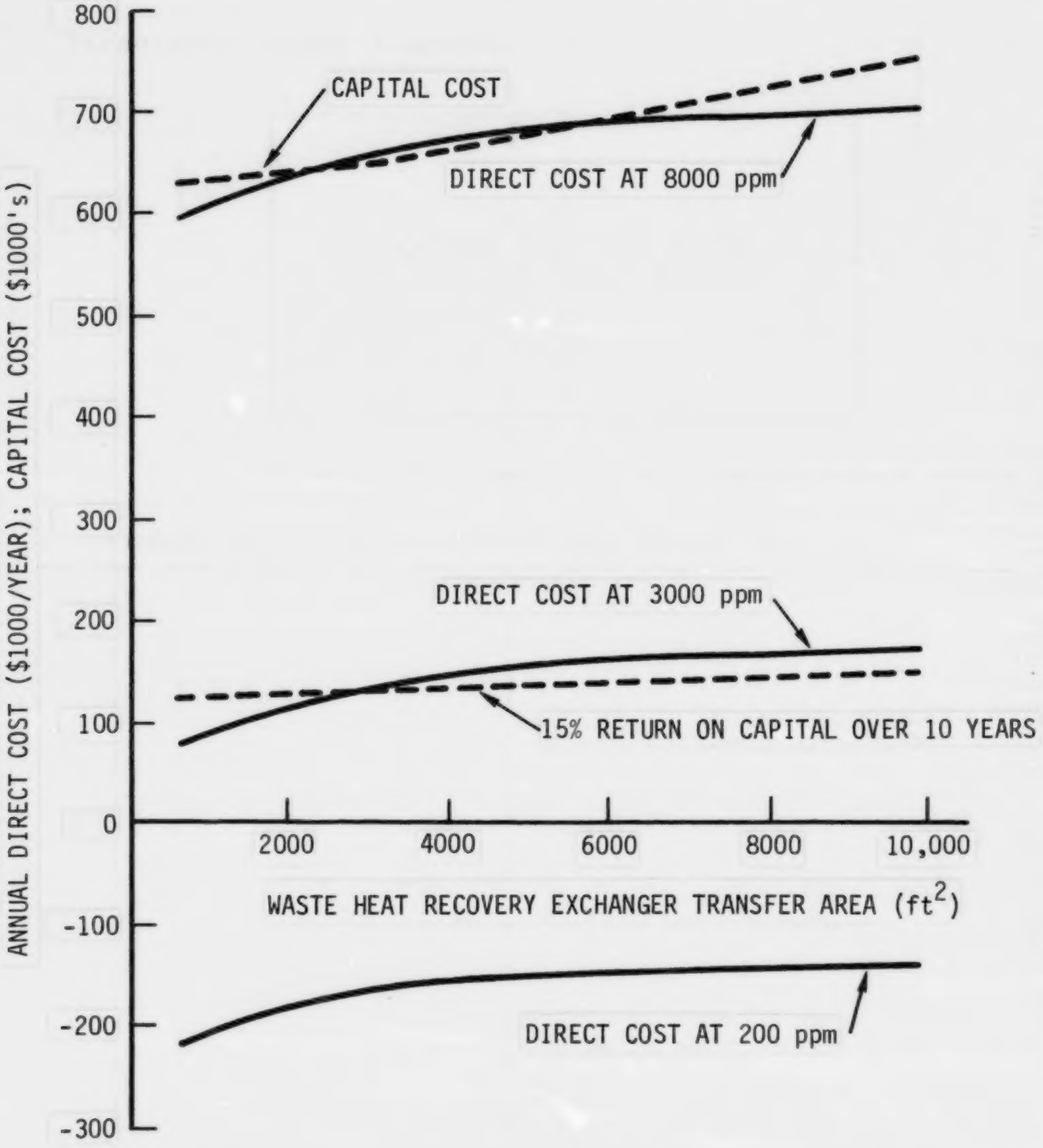

Figure 22. Annual direct costs versus waste heat exchanger size and stack gas concentration.

For 8000 ppm stack gas the highest ratio of direct cost to capital cost is approximately 1.3 when the exchanger heat transfer area is about 4000 ft^2. The least difference between required return for 15 percent interest and the direct costs with 200 ppm stack gas is for an exchanger area of 6000 ft^2. Obviously, a refrigeration/condensation system would not be recommended for very low mean stack gas concentrations. The costs tabulated below are for the 200 ppm detailed system design with optimum capacity exchanger.

8.1.4 System Costs

Capital Cost--	
1980 packed tower cost, installed (8)	$ 27,500
1980 refrigeration system cost, installed (18)	278,200
Decanter cost	10,000
Precooler and waste heat recovery exchangers cost	156,100
Exchangers installation	62,400
Total installed equipment	534,200
Capital cost adjustment	157,400
Total	$691,600
Operating Cost--	
Maintenance cost (10 percent of installed equipment)	$ 53,400
Power cost	137,600
Labor cost (0.5 man years)	9,000
Total	$200,000/year
Operating Savings--	
Solvent reused	
Solvent value	$ 21,000
Waste heat recovered value	31,400
Total	$ 52,400/year
Solvent as fuel	
Solvent value	$ 6,700
Waste heat recovered value	31,400
Total	$ 38,100/year

The preceding refrigeration/condensation system was sized to accommodate a range of stack gas concentrations including the lower limit of 200 ppm. If the minimum expected concentration is greater than 200 ppm, capital costs will be lower. A summary of revised costs for 200, 3000, and 8000 ppm design concentrations are presented in Table 31. The costs presented are for systems with maximum practical precooler capacity and waste heat recovery exchangers sized to maximize the ratio of earned revenue (positive direct cost) to capital cost or to minimize the difference between the annual required return when capital is borrowed at 15 percent interest over 10 years and the annual expenditures (negative direct costs).

8.2 DETAILED DESIGN OF THERMAL INCINERATION SYSTEM WITH HEAT RECOVERY

8.2.1 System Configuration

It was established in the preliminary analysis that the heating value of the VOCs in 200 ppm stack gas is insufficient to achieve the minimum incineration temperature of 1200°F. Supplementary fuel is therefore required for thermal incineration control of emissions at low concentrations. The amount of fuel required can be reduced by preheating the stack gas with the incinerator exhaust gas. Additional energy may be recovered from the incinerator exhaust if the incinerator exhaust temperature leaving the preheater is high enough. The following analysis addresses the system configuration illustrated in Figure 23.

8.2.2 Detailed System Design (200 ppm Case)

Incineration Process--

For the 200 ppm case being considered, the mass flow rates of the inlet stack gas components are:

Component	$\dot{m}$ (lb/min)
N_2	214.0
O_2	64.6
Water	64.1
Solvent	0.37
Fuel	$\dot{m}_f$

During incineration the fuel and solvent combine with oxygen from the air to form combustion products. Assuming that natural gas, which is predominantly methane, is the supplementary fuel, the following chemical equations apply:

$$CH_4 + 2\ O_2 \rightarrow CO_2 + 2\ H_2O \qquad (30)$$

$$C_{10}H_{20} + 15\ O_2 \rightarrow 10\ CO_2 + 10\ H_2O \qquad (31)$$

Hence 1 mole of CH_4 consumes 2 moles of O_2 and produces 1 mole of CO_2 and 2 moles of H_2O. Given 95 percent efficient combustion of $\dot{m}_f$ lb/min of CH_4, the rates of O_2 consumption and CO_2 and H_2O generation due to combustion of the fuel gas are determined to be $3.8 \times \dot{m}_f$ lb/min, $2.61 \times \dot{m}_f$ lb/min, and $2.14 \times \dot{m}_f$ lb/min, respectively. A similar analysis of 0.37 lb/min solvent combustion yields 1.21 lb/min O_2 consumed, 1.10 lb/min CO_2 generated and 0.45 lb/min H_2O generated. The combustion product mass flow rates are then:

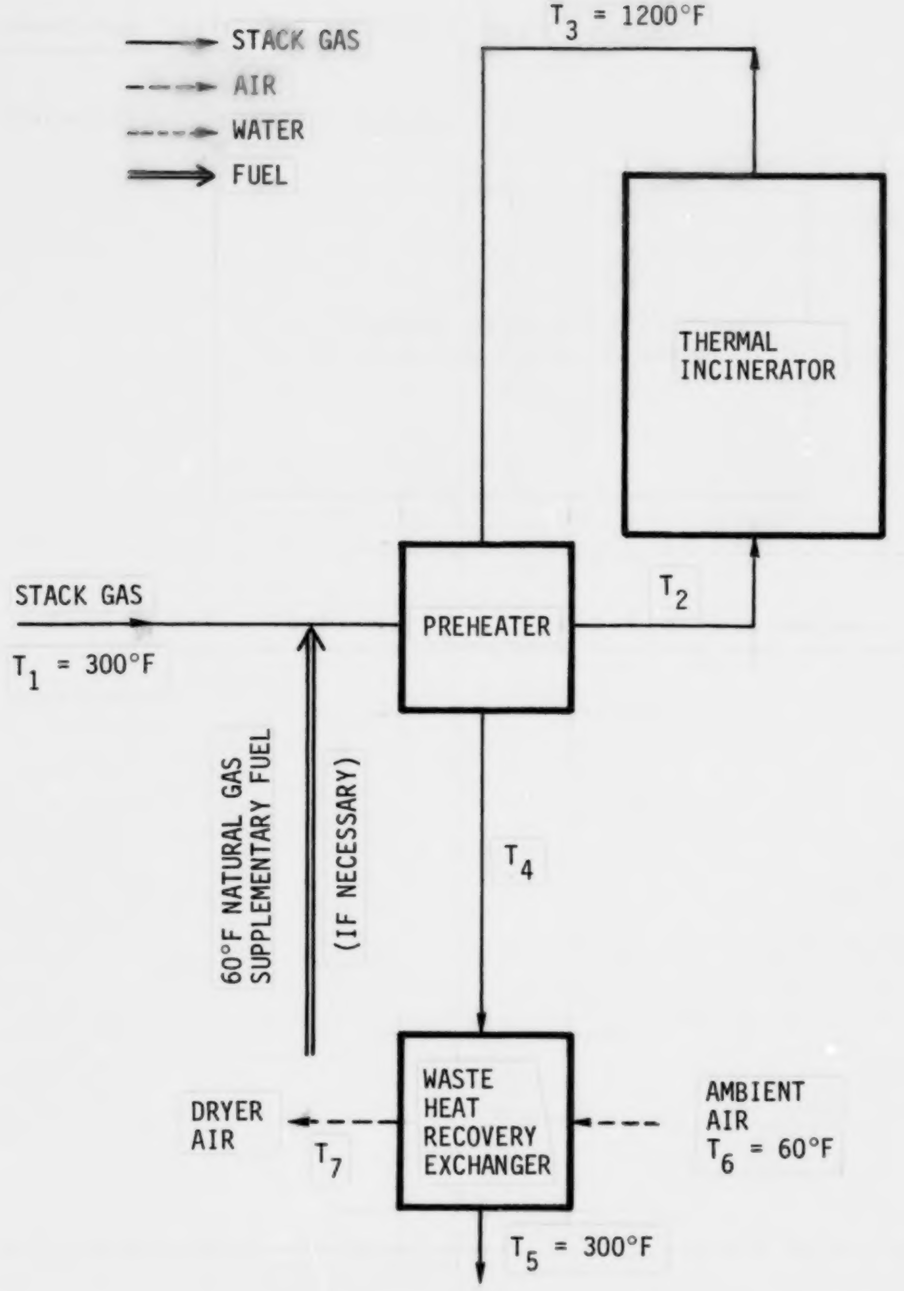
STACK GAS
AIR
WATER
FUEL
T3 = 1200°F
THERMAL INCINERATOR
STACK GAS
T1 = 300°F
PREHEATER
T2
60°F NATURAL GAS SUPPLEMENTARY FUEL
(IF NECESSARY)
T4
DRYER AIR
T7
WASTE HEAT RECOVERY EXCHANGER
AMBIENT AIR
T6 = 60°F
T5 = 300°F

Figure 23. Thermal incineration system configuration for low emissions concentrations.

Component	$\dot{m}$ (lb/min)
N_2	214.0
O_2	$63.39 - 3.80 \times \dot{m}_f$
Water	$64.55 + 2.14 \times \dot{m}_f$
CO_2	$1.10 + 2.61 \times \dot{m}_f$
Fuel	$0.05 \times \dot{m}_f$
Solvent	0.02

The thermal demand of the incinerator is supplied by 95 percent efficient combustion of the solvent and fuel in the stack gas. This thermal balance is expressed by:

$$\dot{m}\,\bar{c}_p\,(T_3 - T_2) = 0.95 \times (\dot{m}_s \times h_{v,s} + \dot{m}_f \times h_{v,f}) \tag{32}$$

where

$\bar{c}_p$ = mean specific heat of mixture in incinerator, Btu/lb°F

$h_{v,s}$ = heating value of solvent, Btu/lb

$h_{v,f}$ = heating value of fuel, Btu/lb.

With $\dot{m} = 343.1 + \dot{m}_f$, the mass flow rate of supplementary fuel required for a 1200°F incineration temperature is

$$\dot{m}_f = \frac{343.1 \times \bar{c}_p \times (1200 - T_2) - 0.95 \times 0.37 \times 19{,}930}{0.95 \times 21{,}270 - \bar{c}_p \times (1200 - T_2)} \tag{33}$$

Assume the incinerator inlet temperature is $T_2 = 1000$°F. Then,

$$\dot{m}_f = \frac{68{,}620\,\bar{c}_p - 7005.4}{20206.5 - 200. \times \bar{c}_p} \tag{34}$$

For a first approximation of $\dot{m}_f$, $\bar{c}_p$ is estimated from

$$\bar{c}_p = \frac{\left(\underset{\substack{\text{inlet components}\\ \text{at 1000°F}}}{\sum \dot{m}\,c_p} + \underset{\substack{\text{combustion products}\\ \text{at 1200°F}}}{\sum \dot{m}\,c_p}\right)}{2.0 \times \sum \dot{m}} \tag{35}$$

using the appropriate component specified heat values from Table 30 and neglecting the contribution of $\dot{m}_f$. With $\bar{c}_p$ estimated to be 0.320 Btu/lb°F the first approximation of $\dot{m}_f$ is 0.74 lb/min. Repeating the evaluation of $\bar{c}_p$ including the effect of $\dot{m}_f$ yields $\bar{c}_p = 0.321$ Btu/lb°F and finally $\dot{m}_f = 0.75$ lb/min.

Preheater--

The thermal load of the preheater is

$$H_{\text{preheater}} = \sum_{\substack{\text{stack gas}\\ \text{components}}} \dot{m}\,\bar{c}_p\,(T_2 - T_1) \tag{36}$$

with $\bar{c}_p$ in this case being the mean component specific heat in the preheater. If natural gas at 60°F is added at the preheater inlet for supplementary fuel,

$$\begin{aligned} H_{\text{preheater}} &= (214.0 \times 0.260 + 64.6 \times 0.238 + 64.1 \\ &\quad \times 0.484 + 0.37 \times 0.641) \times (1000 - 300) \\ &\quad + 0.74 \times 0.772 \times (1000 - 60) \\ &= 72{,}130 \text{ Btu/min.} \end{aligned}$$

Assuming a 10 percent heat loss during stack gas preheat, the following expression for the rate of heat extraction from the combustion products:

$$H_{\text{preheater}} = 0.90 \sum_{\substack{\text{incinerator}\\ \text{exhaust}\\ \text{components}}} \dot{m}\,\bar{c}_p\,(T_3 - T_4) \tag{37}$$

can be used to determine the temperature of the exhaust products as they leave the preheater, T_4. That is,

$$T_4 = T_3 - \frac{H_{\text{preheater}}}{0.90 \times \sum\limits_{\substack{\text{incinerator}\\ \text{exhaust}\\ \text{components}}} \dot{m}\,\bar{c}_p} \tag{38}$$

As $\bar{c}_p$ is the mean specific heat of the incinerator exhaust components in the preheater, it is a function of T_4. The value of T_4 is therefore derived from Equation 38 by iteration. For the example being considered T_4 is first estimated to be 400°F. Using the appropriate value of $\bar{c}_p$ in Equation 38:

TABLE 30. SPECIFIC HEATS (Btu/lb°F) OF STACK GAS COMPONENTS VERSUS TEMPERATURE (°F)

Stack gas component	Temperature (°F)									
	300	400	500	600	700	800	900	1000	1100	1200
N_2	0.247	0.250	0.254	0.258	0.262	0.266	0.270	0.273	0.276	0.279
O_2	0.224	0.229	0.233	0.237	0.240	0.244	0.247	0.250	0.252	0.254
CO_2	0.228	0.239	0.248	0.257	0.264	0.271	0.276	0.282	0.286	0.290
Water	0.448	0.457	0.467	0.478	0.489	0.500	0.510	0.521	0.531	0.540
Solvent	0.477	0.524	0.571	0.618	0.665	0.712	0.759	0.806	0.852	0.899
Fuel (CH_4)	0.654	0.705	0.756	0.808	0.859	0.910	0.962	1.013	1.064	1.116

$$T_4 = 1200°F - 72,130/[0.90 \times (214.0 \times 0.265 + 60.5 \times 0.243 + 66.2 \times 0.499 + 3.06 \times 0.269 + 0.0375 \times 0.910 + 0.02 \times 0.712)$$

$$= 439°F$$

Repeating the procedure followed above for several values of T_2 from the range of possible values, the variation of stream temperatures with supplementary fuel usage rates can be determined. The results of this process are presented in Figure 24.

Waste Heat Recovery--

Since there is still recoverable thermal energy available in the incinerator exhaust gas when it exits the preheater, waste heat recovery exchangers can be used to warm ambient air to $T_7 \leq 400°F$ for the textile dryers and if sufficient heat is left, to preheat boiler feedwater. As in previous analyses, the minimum waste heat recovery gas exhaust temperature is 300°F to prevent condensation of corrosive combustion products. The energy recovered by the waste heat exchangers, again assuming a 10 percent heat loss, is in general

$$H_{WH} = 0.90\ \dot{m}\ \bar{c}_p\ (T_4 - 300) \tag{39}$$

and for this example,

$$H_{WH} = 0.90 \times 343.8 \times 0.285 \times 139$$

$$= 12,260 \text{ Btu/min}$$

Let the dryer air flow rate through the primary waste heat recovery exchanger be equal to adjust air flow rate to the incinerator. For 200 ppm, the dry air flow rate is $\dot{m}_a = 278.6$ lb/min, and

$$T_7 = T_6 + \frac{H_{WH}}{\dot{m}_a\ \bar{c}_{p,a}} \tag{40}$$

Again following through with our example,

$$T_7 = 60 + \frac{12,260}{278.6 \times 0.241}$$

$$= 243°F$$

If lower values of preheat temperature, T_2, are initially assumed, the heat recoverable from the preheater exhaust H_{WH} will be greater. For $H_{WH} > 22,830$ Btu/min, a secondary heat exchanger which heats 100 gpm of 60°F boiler feedwater with the primary heat exchanger exhaust is assumed.

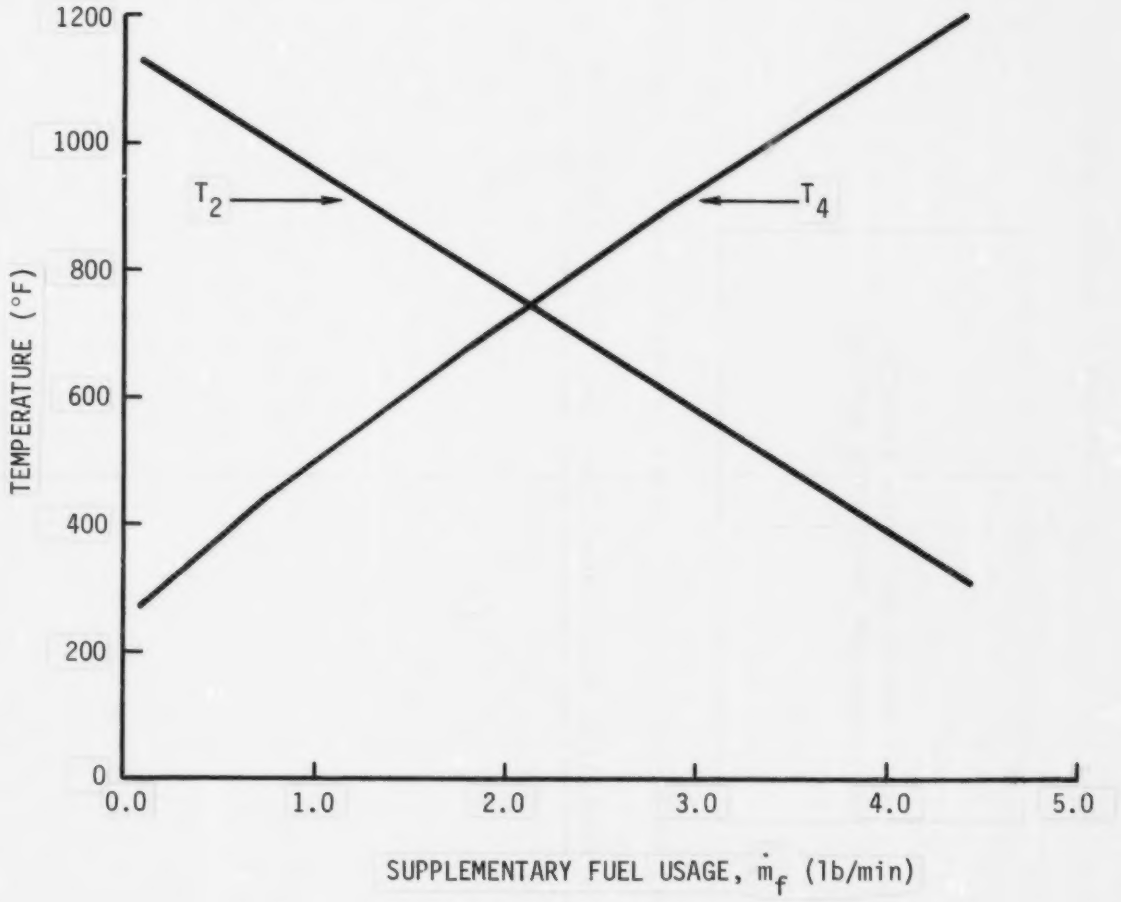

Figure 24. Variation of gas stream temperatures with supplemental fuel usage for thermal incineration with preheat.

Sizing of System Heat Exchangers--

The size of a simple counterflow heat exchanger is related to its thermal load by

$$H = U \times A \times LMTD/60 \tag{41}$$

where a reasonable value for the overall heat transfer coefficient, U, of a gas-to-gas exchanger is 3.5 Btu/hr (ft^2) (°F) and LMTD is the log mean temperature difference between the two gas streams in the exchanger. Therefore, from Equation 41 the preheater heat exchange area is found to be:

$$A_{preheater} = \frac{72,130 \times 60}{3.5 \times 168} = 7360\ ft^2$$

and the waste heat recovery exchanger area is:

$$A_{WH} = \frac{12,260 \times 60}{3.5 \times 217} = 969\ ft^2$$

Sizing of Thermal Incinerator--

The size and therefore cost of a thermal incinerator depends on the actual incinerator exhaust volume flow rate. While the exhaust volume flow rate is primarily a function of temperature according to Equation 17, it also increases slightly due to combustion of the solvent as is implied by chemical Equations 30 and 31. Each mole of $C_{10}H_{20}$ incinerated results in four additional moles of combustion product. Since 0.37 lb/min solvent incinerated at 95 percent efficiency is 0.251 moles/min burned, the exhaust flow rate is 1.00 moles/min at 379.4 standard cubic feet per mole or 381 scfm higher than the inlet flow rate. The exhaust flow rate for 200 ppm stack gas incineration at 1200°F is therefore found to be 17,330 ± 100 actual ft^3/min for $400 < T_2 < 1000$°F. The variation of incinerator exhaust flow rate with T_2 is simply the result of varying supplementary fuel flow rates. Since the variation is so slight, incinerator size and therefore cost are assumed not to vary with T_2.

The maximum permissible incinerator velocity is 20 ft/sec which requires the incinerator diameter to be greater than 4.3 ft. If a 5 ft diam chamber is selected and the recommended incinerator length to diameter ratio of 2 is used then the actual incinerator velocity will be 14.7 ft/sec, the incinerator length will be 10 ft, and the retention time of gases in the incinerator will be 0.68 sec which is consistent with the recommended value of 0.6 sec.

8.2.3 Thermal Incineration System Optimization

The cost of a thermal incineration system for 200 ppm stack gas will vary with amount of stack gas preheat employed.

The capital cost components as a function of preheat temperature, T_2, are presented in Figure 25. The cost of the preheater increases rapidly with T_2 since decreasing mean temperature difference and increasing heat transfer rate combine to accelerate the requirement for heat exchanger area. The dryer air heater increases in size and therefore cost as T_2 increases because the gas exiting the preheater from which the heat is extracted decreases in temperature. The boiler feedwater heater decreases in size since it basically removes whatever heat has not been extracted by the two previous exchangers.

Annual charges for power and labor are essentially constant. The cost of maintenance is assumed to be proportional to installed equipment cost and increases with increasing preheat temperature. As Figure 24 illustrates, the rate of supplementary fuel usage, and therefore, annual cost of fuel decreases with increasing T_2.

Annual savings due to recovery of waste heat from the incinerator exhaust as it exits the preheater also decreases with increasing values of T_2, since the amount of heat extracted by the preheater increases. The difference between savings and operating costs or direct cost remains essentially the same with increasing T_2. This is true because the rate of decrease in fuel plus maintenance costs is matched by the rate of decrease in operating savings. Figure 26 illustrates the variation of annual operating costs and savings with incinerator preheat temperature, T_2.

Optimization of thermal incineration for low concentration stack gas involves minimizing the annual cost of the system including cost of capital. The best configuration for thermal incinerator of low concentration stack gas is assumed to be that for which the difference between annual required return on capital at 15 percent interest over 10 years and the annual expenditures (negative direct cost) is least. This occurs when no preheater is used. Thus the fuel savings resulting from preheater use are not sufficient to justify the capital investment required.

As was determined in the preliminary design of thermal incineration systems, supplementary fuel is required as long as the stack gas concentration is 2600 ppm or less. As the stack gas concentration increases from 200 to 2600 ppm, system capital cost and operating savings will remain essentially constant while the cost of fuel will decrease. Beyond 1250 ppm operating savings will exceed operating costs so that the direct cost will be positive and revenue will be generated.

The system costs listed here are for a thermal incineration system designed for 200 ppm stack gas. Waste heat recovery exchangers are employed, but no preheater is used.

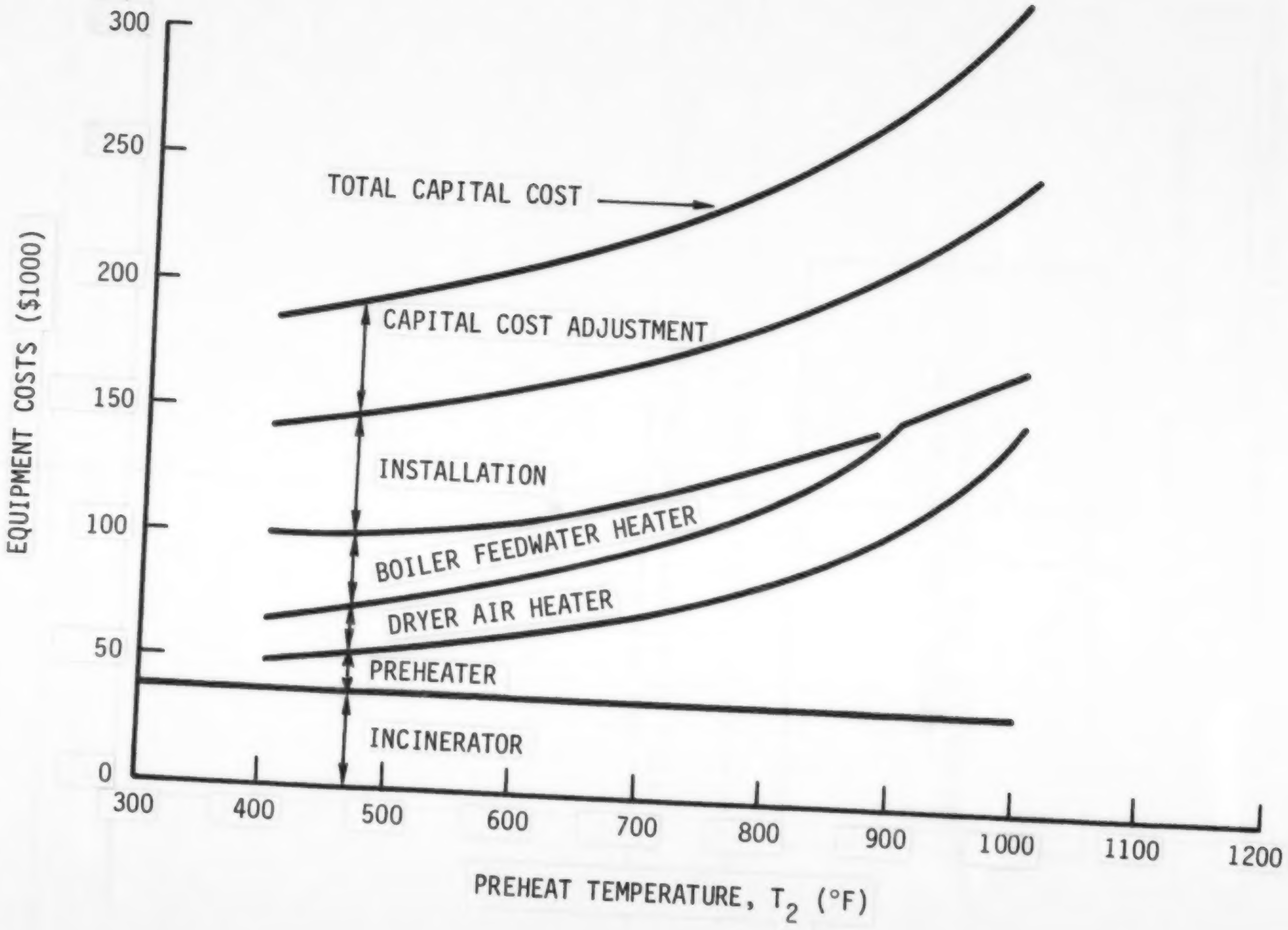

Figure 25. Variation of capital cost components with preheat temperatures.

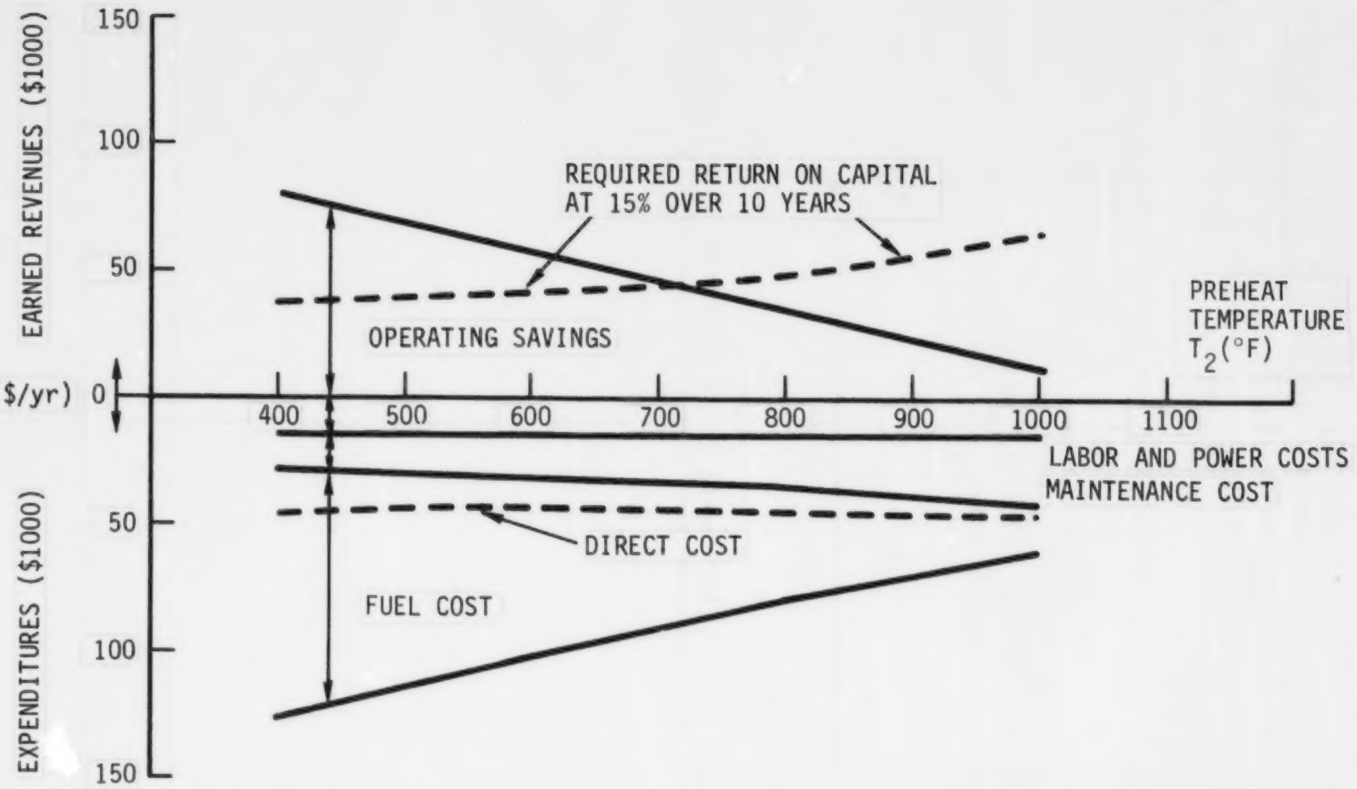

Figure 26. Variation of annual costs and savings with preheat temperature.

8.2.4 System Costs

Capital Cost--

1980 incinerator cost (9)	$ 38,600
Incinerator installation (9)	19,300
Waste heat recovery exchangers	52,200
Exchanger installation (9)	20,900
Total installed equipment	131,000
Capital cost adjustment	38,600
Total	$169,600

Operating Cost--

Maintenance cost (10 percent of installed equipment)	$ 17,000
Labor (0.75 man years)	13,500
Fuel cost	98,600
Power cost	2,000
Total	$131,100/year

Operating Savings--

Dryer air heating	$ 23,900
Boiler feedwater heating	67,200
Total	$ 91,100/year

A summary of costs for thermal incineration systems with heat recovery are presented in Table 31. The costs for the 3000 and 8000 ppm system are as derived in the preliminary system analysis with a 10 percent reduction in operating savings to account for thermal losses in the waste heat recovery process.

8.3 CONCLUSIONS BASED ON DETAILED SYSTEM DESIGNS

Table 31 summarizes the capital costs and annual direct costs of emissions control technologies which generate revenue at stack gas concentrations of 8000 ppm or less for the given stack gas temperature, flow rate and relative humidity. Each system has been designed for the specified emissions concentration and a 10 percent thermal loss has been assumed for each heat exchanger employed. If a heat recovery exchanger is used to warm dryer air, the mass flow rates on each side of the exchanger have been assumed equal. If the recovered waste heat is transferred to boiler feedwater a flow rate of 100 gpm has been assumed. As was the case for all previous cost reports, annual direct cost is equal to annual operating savings minus annual operating costs. Therefore, a positive direct cost indicates earned revenue and a negative direct cost indicates a net annual expenditure.

A more comprehensive economic measure than payback is, of course, required to determine the optimum emissions control technology to incorporate into a textile processing system. The viability of an emissions control system will be influenced

TABLE 31. SUMMARY OF COSTS FOR SELECTED EMISSIONS CONTROL TECHNOLOGIES INCLUDING DETAILED DESIGNS

Stack gas emissions concentration (ppm, volume)	Control technology Resource recovery	Refrigeration/ condensation, aerosol removal		Thermal incineration (stack gas diluted to 3500 ppm, if higher)	Fixed bed catalytic incineration (stack gas diluted to 1580 ppm, if higher)	Fluidized bed catalytic incineration	Fixed bed carbon adsorption (stack gas diluted to 3500 ppm, if higher)		Fluidized bed carbon adsorption	
	Cost item	Recovered solvent reused	Recovered solvent used as fuel	Heat recovery	Heat recovery	Heat recovery	Recovered solvent reused	Recovered solvent used as fuel	Recovered solvent reused	Recovered solvent used as fuel
200	Total capital cost ($)	691,600	691,600	169,600	185,700	214,300	140,000	140,000	84,800	84,800
	Annual direct cost ($/year)	-147,600	-161,900	-40,000	-40,000	-43,700	-9,700	-24,700	+10,500	-4,500
3000	Total capital cost ($)	706,300	706,300	174,800	240,400	224,800	140,000	140,000	122,500	122,500
	Annual direct cost ($/year)	+175,800	-35,200	+69,600	+49,200	+47,100	+228,600	+21,800	+280,900	+74,100
8000	Total capital cost ($)	613,500	685,700	324,100	469,300	256,500	259,600	259,600	177,200	177,200
	Annual direct cost ($/year)	+687,200	+160,000	+219,900	+182,300	+207,900	+643,800	+91,800	+767,400	+215,400

Notes: Negative annual direct cost indicates expenditure; positive annual direct cost indicates earned revenues.

by a list of economic and plant parameters including escalation rates for operating and maintenance costs, taxes and tax incentives, interest rate on money for capital investments, rate of general inflation, etc. A detailed economic evaluation of the selected systems is beyond the scope of this report. However, in order to establish which systems have the greatest potential for application, an annual return rate for each of the systems and stack gas concentrations has been determined using the following model of capital recovery in a uniform series:

$$R = C \frac{i\,(1 + i)^n}{(1 + i)^n - 1} \tag{42}$$

where

R = annual return required on capital investment, dollars/year

C = capital cost of system, dollars

i = annual rate of return, fraction

n = system life, years

Thus, the rate of return values, i, associated with the revenue generated, R (positive direct cost), and money invested, C (capital cost), for each system as reported in Table 31 have been determined and are presented in Table 32. A 10 year system life has been assumed.

The estimates of rate of return on investment (ROI) presented in Table 32 show that:

- Fluidized bed carbon adsorption with solvent reused projects the best ROI at every stack gas concentration level by a wide margin. It is also the only system which generates revenue at 200 ppm. Fluidized bed carbon adsorption is still the best technology overall if recovered solvent can only be used as fuel.

- Fixed bed carbon adsorption with solvent reused projects excellent ROI values for 3000 and 8000 ppm. If recovered, solvent can only be used as fuel; however, the rate of return is seriously reduced so that other technologies are preferable.

- The refrigeration/condensation aerosol removal system is competitive only at 8000 ppm stack gas concentration and only if the recovered solvent can be reused.

TABLE 32. ESTIMATES OF RATE OF RETURN ON CAPITAL INVESTMENT FOR SELECTED EMISSIONS CONTROL TECHNOLOGIES INCLUDING DETAILED DESIGN COSTS

Control technology	Refrigeration/ condensation, aerosol removal		Fixed bed carbon adsorption (stack gas diluted to 3500 ppm, if higher)		Fluidized bed carbon adsorption		Thermal incineration (stack gas diluted to 3500 ppm, if higher)	Fixed bed catalytic incineration (stack gas diluted to 1580 ppm, if higher)	Fluidized bed catalytic incineration
Stack gas emissions concentration (ppm, volume) / Resource recovery	Recovered solvent reused	Recovered solvent used as fuel	Recovered solvent reused	Recovered solvent used as fuel	Recovered solvent reused	Recovered solvent used as fuel	Heat recovery	Heat recovery	Heat recovery
200	-	-	-	-	0.04	-	-	-	-
3000	0.21	-	1.63	0.09	2.29	0.60	0.38	0.16	0.16
8000	1.12	0.19	2.48	0.33	4.33	1.22	0.67	0.37	0.81

- Thermal incineration provides reasonable ROI's at 3000 and 8000 ppm.
- Fixed bed and fluidized bed catalytic incineration generate competitive returns only at 8000 ppm.
- At 8000 ppm fluidized bed catalytic incineration is preferable to either thermal incineration or fixed bed catalytic incineration.

The top six technologies ranked by potential return over the range of concentration levels are therefore:

- Fluidized bed carbon adsorption with recovered solvent reused
- Fixed bed carbon adsorption with recovered solvent reused
- Fluidized bed carbon adsorption with recovered solvent used as fuel
- Thermal incineration with heat recovery
- Refrigeration condensation/aerosol removal with recovered solvent reused
- Fluidized bed catalytic incineration with heat recovery.

SECTION 9

EFFECT OF STACK GAS FLOW RATE ON THE ECONOMICS OF SELECTED CONTROL TECHNOLOGIES

The selected control technologies are:

- Fluidized bed carbon adsorption with recovered solvent reused
- Fixed bed carbon adsorption with recovered solvent reused
- Fluidized bed carbon adsorption with recovered solvent used as fuel
- Thermal incineration with heat recovery
- Refrigeration condensation/aerosol removal with solvent reused
- Fluidized bed catalytic incineration with heat recovery.

The size and cost of all the above control devices at a given concentration are a function of the gas flow rate. The equipment size increases in proportion to the gas flow rate. Equipment cost will show scale-up advantages. Of the operating cost items maintenance cost (a percentage of the equipment cost) will show scale-up effects, utility costs are proportional to gas flow rate and labor cost remains essentially constant. The credit for solvent or waste heat recovery is proportional to gas flow rate.

The following are examples of cost variations at 1,000 and 10,000 scfm stack gas flow rate. A scale up index of 0.7 has been used.

EXAMPLE: Fixed bed carbon adsorption with recovered solvent reused, 3000 ppm stack gas concentration, 1000 scfm flow rate.

Total capital cost	$= (1000/5000)^{0.7} \times (140{,}000)$	$ 45,400
Operating costs:		
Steam cost		$ 7,800
Electricity cost		1,400
Cooling water cost		1,800
Labor cost		9,000
Maintenance cost		3,500
Carbon cost		1,300
Total operating cost		$ 24,800
Solvent reuse credit		$ 67,800
Annual direct cost	= 67,800 - 24,800	$+43,000
Return on investment		0.95

EXAMPLE: Fixed bed carbon adsorption with recovered solvent reused, 3000 ppm stack gas concentration, 10,000 scfm flow rate.

Total capital cost	$= (10{,}000/5{,}000)^{0.7} \times (140{,}000)$	$ 227,400
Operating costs:		
Steam cost		$ 78,200
Electricity cost		14,000
Cooling water cost		18,400
Labor cost		9,000
Maintenance cost		17,600
Carbon cost		13,000
Total annual operating cost		$ 150,200
Solvent reuse credit		$ 610,400
Annual direct cost	= 610,400 - 150,200	$+460,200
Return on investment		2.02

If the return on investment for 5,000 scfm is assigned a factor of 1, then the ROI factors for 1,000 and 10,000 scfm operations are 0.58 and 1.24 respectively for 3000 ppm stack gas. Figure 27 shows the ROI factors for the best six technologies versus stack gas flow rate. Hence, if the ROI for one of the selected emissions control systems operating at 3000 ppm is known for a particular flow rate, the ROI for smaller or bigger units can be ascertained from Figure 27.

For the selected systems operating with 8000 ppm stack gas the ROI factors for gas flows higher than 5000 scfm are the same as those shown in Figure 27, except for the fluidized bed catalytic incineration and refrigeration condensation systems. The ROI factors for these two systems are approximately 80 percent of those shown. For 8000 ppm stack gas and a flow rate of 1000 scfm the ROI factors for fluidized and fixed bed adsorptions systems are the same as those illustrated. The ROI factor for all other systems at 8000 ppm and 1000 scfm is approximately 0.49 (±0.05).

With 200 ppm stack gas only the fluidized bed carbon adsorption system with solvent reused projects a return and only for flow rates greater than 5000 scfm.

The ranking of the top three technologies does not vary with stack gas flow rate.

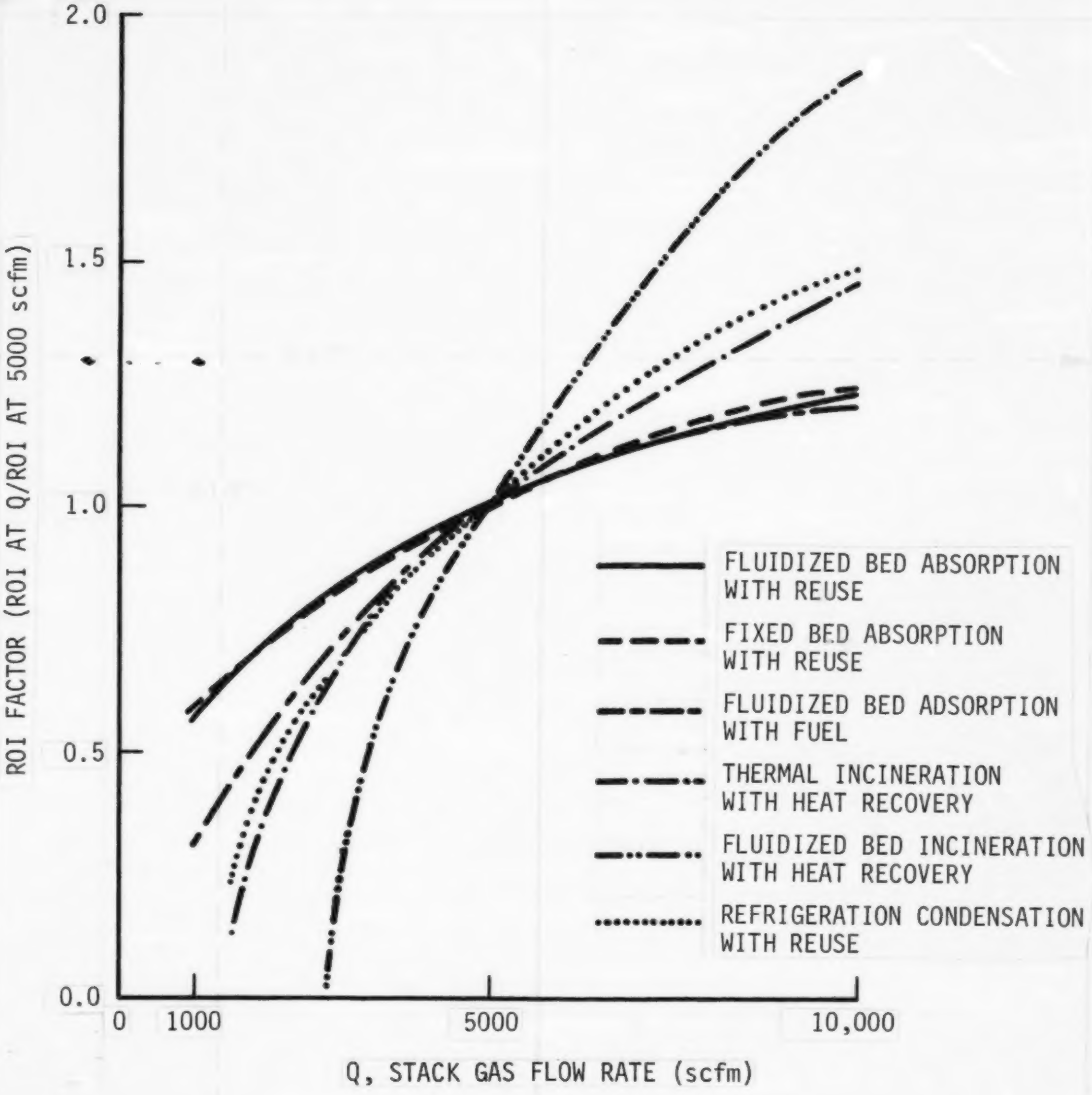

Figure 27. Economics of size for selected emissions control systems.

TYPING GUIDE SHEET

SECTION 10

REFERENCES

1. "Screening Study on the Justification of Developing New Source Performance Standards for Various Textile Processing Operations," prepared for EPA under Contract No. 68-02-0607-11 by Research Triangle Institute, August 1974.

2. "Source Assessment: Overview and Prioritization of Emissions from Textile Manufacturing," prepared for EPA by the Monsanto Research Corporation; EPA-600/2-77-107h, September 1977.

3. "Prioritization of Emissions from Textile Manufacturing Operations on the Basis of Potential Impact of New Source Performance Standards," prepared for EPA under Contract No. 68-02-2612-64 by Research Triangle Institute, April 1979.

4. "Final Report, Phase I: Textile Printing", Emission Standards and Engineering Division, Chemicals and Petroleum Branch, EPA, Office of Air and Waste Management, Office of Air Quality Planning and Standards, October 1979 (as yet unpublished).

5. "Air Quality Criteria for Petrochemical Oxidants," U.S. Department of Health, Education and Welfare, AP-63, March 1970.

6. Information provided by Department of Natural Resources of the States of Georgia and North Carolina, and the Air Pollution Management Department of the State of Rhode Island.

7. "Adsorption Systems," Union Carbide Corporation, Reprinted Brochure in Chemical Engineering, 1973.

8. Air Pollution Engineering Manual, Report No. 999-AP-40, U.S. Department of Health, Education and Welfare, 1967.

9. "Capital and Operating Costs of Selected Air Pollution, Control Systems," GARD, Inc., EPA Report No. EPA-450/3-76-014, May 1976.

10. "Activated Carbon Methods for Controlling Stationary Source Air Pollutants," Calgon Corporation, Adsorption Conference, Pittsburgh, PA, 1975.

11. Heat Recovery Equipment for Exhaust Gases, Oak Ridge National Laboratory, ANL/CES/TE-77-4, April 1977.

12. Advanced Coal Fueled Combustor/Heat Exchanger Technology Study, Heat Exchanger Design Report under Contract No. EF-77-C-01-2612, Rockwell International Corporation for the Department of Energy, June 1979.

13. Telephone conversations with industrial oil and gas suppliers, 1980.

14. Telephone information obtained from the Electric Power Research Institute, Palo Alto, CA, 1980.

15. Telephone information from manufacturers, 1980.

16. Githens, R.E. and D.M. Sowards, "Catalytic Oxidation of Hydrocarbon Fumes, " Proceedings: Textile Industry Technology Symposium, Williamsburg, VA, December 1978.

17. Chandrasekhar, R. and C.M. Yon, "Fluidized Bed Activated Carbon Adsorption," Ibid.

18. Perry's Chemical Engineering Handbook, Fifth Edition, McGraw-Hill Publications, 1976.

19. Herbank, W.H., "Incineration and Heat Recovery in the Textile Industry," paper presented at the Textile Industry Technology Symposium, Williamsburg, VA, December 1978.

20. Private communications, L.C. Hardison, Air Resources, Inc. Palatine, IL.

21. Glicksman, L.R. and N.A. Decker, Proceedings of the 6th International Conference on Fluidized Bed Combustion, April 1980, Atlanta, GA.

22. Sakaguchi, Y., "Development of Solvent Recovery Technology using Activated Carbon," Chemical Economy and Engineering Review, December 1976.

23. Direct Contact Heat Exchangers for Geothermal Application, DSS Engineers, Inc. for Lawrence Livermore Laboratories, Contract No. E(40-1)4893, 1977.

APPENDIX A

BOOKS, JOURNALS AND OTHER PUBLICATIONS CONSULTED IN LITERATURE SURVEY

A list of publications consulted in the course of the literature survey follows.

BOOKS

Modern Pollution Control Technology, Staff of Research and Education Association, Copyright 1978 by Research and Education Association

Pollution Control Technology, Staff of Research and Education Association, Copyright 1973 by Research and Education Association

Industrial Pollution Control Handbook, Herbert F. Lund, McGraw-Hill Inc., 1971

Air Pollution, Henry C. Perkins, McGraw-Hill Inc., 1974

JOURNALS

J. Air Pollution Control Association
J. American Association of Chemists and Colorists
American Dye Stuff Reporter
Modern Textiles
Textile World
Textile Industries
Textile Products and Processes
Chemical Engineering
Chemical Engineering Progress
Chemical Weekly
Chemical Processing
Pollution Engineering
Environmental Science and Technology
Atmospheric Environments

OTHER PUBLICATIONS

Proceedings, Hydrocarbon Control Feasibility: Its Impact on Air Quality - Specialty Conference: Air Pollution Control Association, April 1977 at Cooper Union, N.Y.C., N.Y.

Control Technology: Gases & Odors, Air Pollution Control Association, August 1973, Library of Congress Catalogue Card 73-87988

Control Technology Evaluation: Gasoline Loading Barges, Pullman-Kellog, Houston, TX, March 1979, PB-298016

Control of VOC Emissions from Perchloroethylene Dry Cleaning Systems, OAQPS No. 1.2-117, PB-290613, December 1978

Control of VOC Emissions from Manufacture of Synthetic Pharmaceuticals, OAQPS No. 1.2-105, PB-290580, December 1978

Control of Hydrocarbon and Carbon Monoxide Emissions from Coal Gasification Facilities, for the Division of Environmental Control Technology of DOE, by Oakridge National Laboratories, ORNL/TM-6229, August 1978

Control Technology: Particulates, Air Pollution Control Association, July 1973, Library of Congress Catalog Card 73-87081

Particulates and Fine Dust Removal Equipment, Marshall Sitting, Noyes Data Corporation, Park Ridge, NJ, 1977

APPENDIX B

SAMPLE QUESTIONNAIRE

TYPING GUIDE SHEET

Hydrocarbon Emissions from Textile Plants
EPA Contract No. 68-02-3134
Contractor: Foster-Miller Associates, Inc.

Company Name: ______________________________

Plant Name: ______________________ *Number: __________

Location: ______________________________

Prepared by: Name: ______________________

Title: ______________ Phone No. ______

Signature: ______________________

Date: ______________________

*If several plants treat approximately same quantities of similar materials, enter the number of such plants and details for one plant only, otherwise please fill out one form for each group of similar plants.

1. Check Materials Handled

   Cotton ☐ Synthetic ☐ Both ☐

2. Check Operations

   Dyeing ☐ Printing ☐ Finishing ☐

3. Check Other Operations

   Bleaching ☐ Scouring ☐ Mercerizing ☐

4. Have hydrocarbon emissions from stacks been monitored at this plant?

   Yes ☐
   No ☐
   (If yes, answer all questions from page 2 onwards)

5. Are there Pollution Control devices currently in use?

   Yes ☐
   No ☐
   (If yes, answer all questions from page 3 onwards)

   (If answers to both 4 and 5 are NO, answer all questions from page 5 onwards)

6. *If stacks in the plant have not been analyzed, will you consider having EPA representatives carry out tests on your stacks at EPA expense.*

   Yes ☐ *No* ☐

EMISSIONS DATA (Upstream of Control Devices, if used)

1. Which stacks were monitored with respect to operations (for example, Dryer following printing, heat-set tenter frame; A flow diagram showing location of stacks will be very helpful.

____

____

____

____

2. Were the stacks monitored over long periods of time or spot checked?

____

____

3. If spot checked:

a. What material was being processed? ______________________________

b. At what rate (yds. per min. of fabric)? ______________________________

c. Processes involved. Heat Setting ☐

Dyeing ☐

Printing ☐

Finishing ☐

d. Provide all available emissions data in the form you currently have it and please include methods or names of outside companies used for monitoring. If possible, include stack location, stack size, gas flow rate (scfm or fan capacity), temperature, solid particulate concentration, liquid particulate concentration, hydrocarbon species identified, their concentrations both in liquid and vapor phases and moisture content.

POLLUTION CONTROL DATA

1. Location of pollution control devices (which stack or stacks). Please provide a flow diagram.

2. Device Types (for example, Scrubber, ESP - Coordinate with response to 1)

3. Downstream operating parameters for each device/stack. (Upstream data should be provided on Page 2)

   a. Were the stacks monitored after the poll-control device over long periods of time or spot checked?

   b. If spot checked:

      What material was being processed?

      At what rate (yds/min)?

      Processes involved: Heat Setting ☐

      Dyeing ☐

      Printing ☐

      Finishing ☐

   c. Provide all available emissions data in the form you currently have it and please include methods or names of outside companies used for monitoring. If possible, include stack location, stack size, gas flow rate (scfm or fan capacity), temperature, solid particulate concentration, liquid particulate concentration, hydrocarbon species identified, their concentrations both in liquid and vapor phases, moisture content.

4. Provide vendor/engineer name for each device, their collection efficiency specifications, actual control efficiency, cost of equipment, cost of installation, operating cost (including additional fuel used, if any), savings (if heat or product recovered). Please use separate sheet.

POLLUTION CONTROL DATA (continued)

5. Operating Schedule for each device.

   a. Continuous (regardless of process)________________

   b. Continuous (with process schedule)________________

   c. Intermittent____________Percent time used_____________

   d. Non operative_________________(reasons)

6. What kind of operational problems, if any, have been encountered with each device?

7. How is captured hydrocarbon disposed of, at what estimated operating cost?

8. Has this plant introduced process changes that directly contribute to reduction in hydrocarbon emissions through stacks? If so give details of changes and estimate reduction in stack emissions.

PRODUCTION DATA

1. 1978 Production Rate of each Fabric (cotton, nylon, polyester, blends, etc.)

	Material	Yards	Pounds
for example:	cotton	100,000	175,000
	50-50 cotton/polyester	66,000	100,000

2. Number of operating days in 1978 ____________________.
Working hours per day during 1978.

3. Use of solvents in 1978

	Solvent (chemical/generic or trade name)	Quantity Used (lbs)*			
		Heat Setting	Dyeing	Printing	Finishing
for example:	Varsol		-	200,000	-
	Kerosene		15,000	87,000	-
	Methyl Stearate		50,000	-	-

*Exclude amounts retained in containers, spills, waste and chemicals used in equipment cleaning.

MATERIALS INVENTORY BY OPERATION

If emissions data (Page 2) and/or pollution control date (Page 3) have been provided, the following questions need not be answered.

1. 1978 operating time for each process (hrs/year)

   Heat Setting_____ Dyeing_____ Printing_____ Finishing_____

2. Processing of each fabric and conditions during 1978

Average Fabric Rate (yds/hr), (lbs/yr)	H,D,P,F*	Solvents Used*** (lbs/yr)

3. Number of stacks and stack gas flow rates with temperatures for one "typical" line of each type of operation.

*	Number of Stacks	Stack Gas (Flow rate**), (Temp°)
H		
D		
P		
F		

*H-Heat Setting D-Dyeing P-Printing F-Finishing

**Indicate if flow rate in scfm or acfm, temperature °F or °C Substitute fan capacity for flow rate, if more convenient.

***Exclude amounts retained in containers, spills, wastes and chemicals used in equipment cleaning.

APPENDIX C

CONTROL DEVICES AND SUPPLIERS

A list of control devices and suppliers consulted through telephone and letter communications in the course of the literature survey follows.

Air Correction Division, UOP
TORVEX, Supported Catalyst, DuPont
Carbon Adsorber, Union Carbide Corporation
Carbon Adsorber, Oxy-Catalyst
Carbon Adsorber, Calgon
Carbon Adsorber, Rohm and Haas
Carbon Adsorber, Vic Manufacturing Company
Carbon Adsorber, Per-Ameg Company
Carbon Adsorber, American Ceca Corporation
Carbon Adsorber, Haveg Industries
Incinerator, Baum Company, Inc.
Incinerator, CE Airpreheater
Incinerator, Oxy-Catalyst
Incinerator, Engelhard
Incinerator, Maxon Corporation
Incinerator, National Airoil Burner Company
Incinerator, C & H Combustion
Incinerator, REECO, Inc.
Incinerator, John Zink Company
Incinerator, Airtech System Inc.
ESP, United Air Specialists, Inc.
ESP, Fluid-Ionics Systems
ESP, Beltram Associates, Inc.
ESP, Eltron Manufacturing, Inc.
ESP, Fiber-Dyne Inc.
ESP, Airtech System Inc.
ESP, Micropul Corporation
ESP, Research Cottrell
ESP, Tepco, Inc.
ESP, United McGill Corporation
ESP, Western Precipitation
Scrubbers, Environeering Inc.
Scrubbers, Pollution Control Systems Corporation
Scrubbers, Chemico Air Pollution Control
Scrubbers, Air Pollution Industries, Inc.
Scrubbers, Fiber-Dyne Inc.
Scrubbers, Corborundum Universal

Scrubbers, Centri Spray Corporation
Scrubbers, Ceilcote Company
Scrubbers, Clarkson Controls & Equipment
Scrubbers, W. W. Sly Manufacturing Company
Scrubbers, Stansteel Corporation
Scrubbers, Beco Engineering Inc.
Scrubbers, Swemco, Inc.
Coalescers/Demisters, Beltran Associates Inc.
Coalescers/Demisters, CVM Corporation
Coalescers/Demisters, Monsanto Enviro-Chem
Coalescers/Demisters, Munters Corporation
Coalescers/Demisters, Anderson IBEC

APPENDIX D

PROPERTIES OF VARSOL AS PER MANUFACTURER

Brand name	Varsol 1
Solvency	
Aniline point, °C (°F)	53 (128)
Kauri-butanol value	39
Solubility parameter	7.8
Composition, volume percent	
Total aromatics	18.0
C_8 and higher aromatics	18.0
Olefins	1.0
Saturates	81.0
Volatility	
Flash point, TCC, °C (°F)	42 (108)
Distillation, °C (°F)	
IBP	156 (313)
10 percent	163 (326)
50 percent	175 (347)
90 percent	189 (372)
Dry point	202 (395)
FBP	206 (403)
Vapor pressure, kPa at 38°C	4.8
Vapor pressure, psia at 100°F	0.7
Evaporation rate, seconds	
25 volume percent	2,200
50 volume percent	5,600
75 volume percent	10,900
95 volume percent	19,200
Odor, bulk	Pass
Odor, residual	Pass
Odor, stability	Pass
General	
Gravity, °API	46.4
Specific gravity at 15.6°/15.6°C	0.795
kg/m^3 (lb/gal)	795 (6.62)
Bromine index	200
Sulphur, ppm	1
Copper corrosion, 3 hr at 100°C (212°F)	1
Copper corrosion, 1/2 hr at BP	2

Brand name	Varsol 1
Toxicological Data	
Inhalation TLV, ppm	100
Acute oral LD_{50} (Rat), g/kg	>5
Acute Dermal LD_{50} (Rabbit), g/kg	>2

APPENDIX E

EQUIPMENT COST INDEX

The equipment cost index used in these studies is shown in Figure E-1.

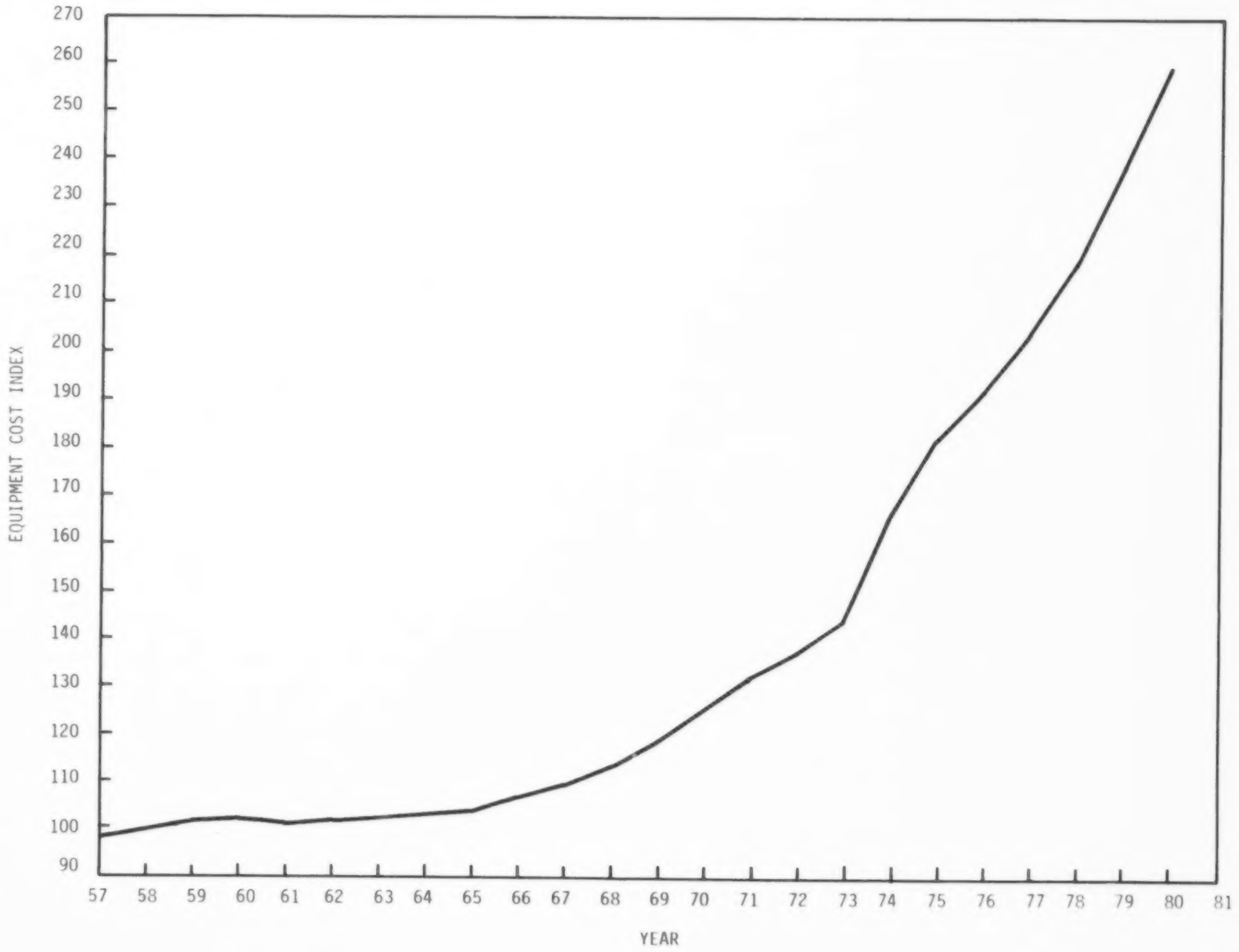

Figure E-1. Chemical engineering equipment cost index.